REDUCING THE MAGNESIUM OXIDE CONTENT IN TROJAN'S NICKEL FINAL CONCENTRATES

MSc (50/50) RESEARCH REPORT

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DECLARATION

I confirm that this is my own unaided work except where I have explicitly indicated otherwise. This work has not been submitted before for any degree or examination to any other University.

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ABSTRACT

Trojan Nickel Mine in Bindura, Zimbabwe, produces nickel concentrates which, until 2008, were then processed at their smelter operations (Bindura Smelter and Refinery) and the subsequent product sent to the hydrometallurgical plant to produce nickel cathodes. However, due to economic challenges the smelter and hydrometallurgical plant operations were closed down in 2008. Currently, Trojan Mine produces nickel concentrates through flotation which are then sold to Glencore International, in China, for further processing.

Since 2002, the MgO (also known as talc) content in the Trojan Nickel Mine final concentrates has increased from around 12% to a peak of 22%. The average MgO content in the concentrates for the year ending in March 2015 was 16.14%. An offtake agreement of sale was made with Glencore International, in China, whereby a penalty is charged for all concentrates with MgO levels greater than 5%. In the year 2015 alone, monthly revenue due to smelter penalties amounted to an estimated total of US$141 000. Higher MgO levels in the concentrates are prevalent when processing low grade ores, with nickel content ranging from 0.65-1.2%. This research focused on reducing the MgO content of the Trojan's final concentrate to 12%; which was the smelter's set target while it was still operational.

In order to investigate the effect of pH and chemical depressants on the MgO levels in the concentrate, batch flotation tests were carried out at pH 8.95 and 10.2, using several guargum depressants namely: Betamin DZT 245 (standard), Cytec S9349, DLM PDE, DLM RS, and CMC (carboxy methyl cellulose) depressants namely: Depramin 177, 267 and 347, and ND 521, 522 and 523. The concentrates were collected at 1, 5, 15 and 25 minute intervals in order to understand the stage-wise recovery of nickel and MgO minerals. A flotation test, without a depressant, was also carried out in order to understand the kinetics of the gangue minerals. Stage addition of depressants was investigated, by adding another 50g/t dose of the DZT 245 depressant after 1 minute into the flotation test. Collector combination tests using SIPX, SIPX:NC228, SIPX:NC236 and SIPX:PNBX, were also carried out to determine the best reagent suite. To understand the recovery of nickel and MgO in the flotation circuit, a plant survey was carried out, and the particle size distribution (PSD) and assays of collected samples were determined.

Flotation tests results indicated that DLM RS and DLM PDE guargum depressants had better selectivity towards MgO and higher nickel recoveries as compared to the Betamin DZT 245 depressant that is currently used in the plant. It is recommended that a plant trial be carried out using the DLM RS depressant, which further reduced the MgO and mass of concentrate recovered by 3.79% and 32% respectively. The stage recovery of MgO for a test carried out without a depressant showed that 57.7% of the MgO was recovered during the first five minutes of the test. Thus, there is need to effectively depress the fast floating MgO during the early stages of the flotation process. Nickel recovery and grade were increased by 2.7% and 2.1% respectively, after adding the second dose of the depressant after 1 minute into the flotation test. The results indicated that the fast floating MgO can depress the valuable mineral if the depressing effect of the depressant is short-lived, which in turn leads to reduced nickel recoveries. Hence, reducing the time between the two stage additions of the depressant in the plant will help further supress the fast floating MgO silicates. It was also noted that at least 60% of the nickel was recovered during the first five minutes of the tests. Hence, reducing the residence time of the rougher flotation bank would reduce MgO recovery into the concentrates without adversely affecting the nickel recoveries.

Plant survey results showed that the scavenger bank feed that was deslimed, had less finer MgO particles and MgO content as compared to the rougher bank feed. This indicates that desliming before the coarse flotation process could reduce MgO slimes in the feed, reduce the recovery of MgO due to slime coatings in the final concentrates and the reagent consumption in the bank. Introducing the desliming unit could be beneficial since the desliming cyclones have low installation and operational costs.

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CHAPTER 1

INTRODUCTION

Most nickel processing plants experience problems with readily floatable magnesia (MgO) bearing gangue minerals present in their ores. Trojan Nickel Mine in Zimbabwe has one of the highest MgO content in its ores. The MgO content in Trojan ores ranges from 28%-42% in disseminated ores which make up the bulk of its ore reserves. There has been a notable increase in the amount of MgO in the nickel concentrates since the company started mining ores with high MgO concentrations in 2002. The MgO content in the concentrates has increased from 12% to around 22%.

Trojan Nickel Mine is owned by Asa Resource group, as well as Bindura Smelter and Refinery division (BSR) that used to process their concentrates and those from other mines in the group. The BSR division is currently not operational. Renovations at the smelter are currently underway and the operations are expected to resume in 2017. The nickel ore is taken through the processing stages that are shown in the flow sheet in Figure 1.1; in order to produce the nickel cathodes.

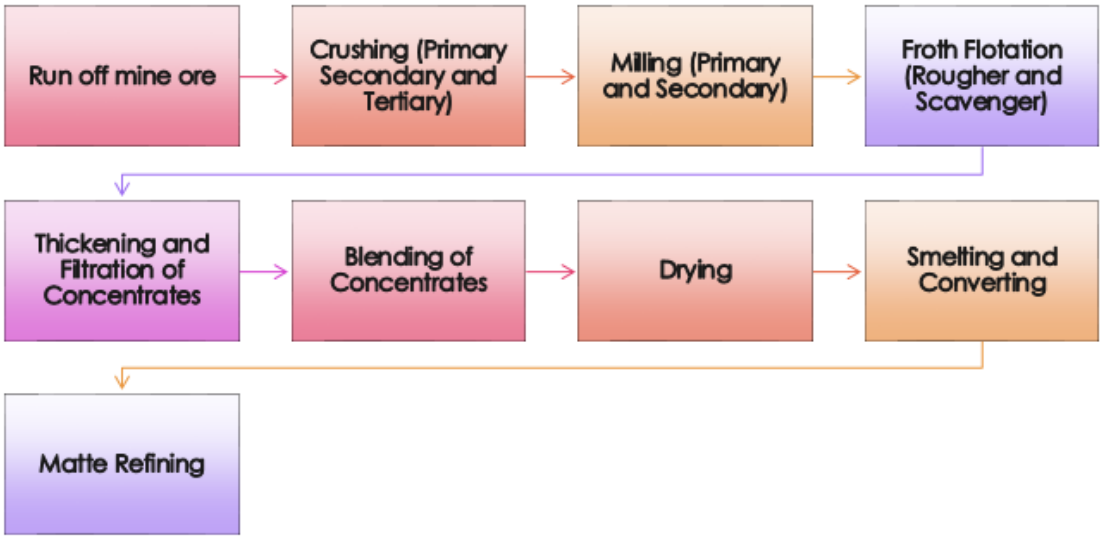

Figure 1.1 Processing stages of the nickel ore to produce nickel cathodes

The concentrates are currently processed up to the drying stage, bagged and shipped to Glencore International, in China.

The concentrator feed usually varies from the low grade feed with an average grade of 0.65% Ni to high grade feed of up to 4% Ni. The high grade feed that is greater than 1.5% Ni consists of low grade ores that are blended with massive ores.

Blending of the ores is carried out so as to lower the MgO content in the mill feed. The blending ratios that are currently employed vary greatly and this results in varying feed compositions ranging from 0.65% to 4% Ni into the plant. Table 1.1; shows the chemical analysis of the concentrator feed as well as the concentrates of the low grade ores and high grade ores. Two assay results have been provided for each category in order to show the variation in the feed and concentrate assays produced in the concentrator plant. From the assay results provided, it can be noted that higher MgO levels in the concentrate are obtained when processing the low grade ores, usually with nickel content ranging from 0.65-1.2%.

Table1.1 Chemical analysis of Trojan's ore feed and concentrates

Content %	Ni	Cu	Co	Fe	SiO2	MgO	CaO	Al_2O_3	S	Cr_2O_3
Low Grade Feed	0.74	0.04	0.01	7.60	39.73	35.80	2.31	2.93		
Concentrate	10.08	0.42	0.22	22.30	27.40	17.79	1.30	1.08	16.48	0.10
Low Grade Feed	0.77	0.04	0.02	7.76	40.30	37.67	1.95	2.72		
Concentrate	9.72	0.36	0.21	20.28	28.32	19.68	0.73	1.19	14.99	0.10
High Grade Feed	2.20	0.10	0.05	13.08	35.47	28.72	2.62	2.97		
Concentrate	16.30	0.90	0.40	22.39	20.94	13.10	0.70	0.87	20.67	0.08
High Grade Feed	2.65	0.14	0.07	14.30	35.28	27.42	2.84	2.99		
Concentrate	14.8	0.77	0.34	25.18	19.68	12.25	0.78	0.96	21.30	0.10
Feed	0.74	0.04	0.01	8.91	41.16	35.12	2.03	2.78		
Off Spec Concentrate	7.97	0.34	0.16	19.31	31.11	21.46	0.71	1.39	12.56	0.10
Feed	0.93	0.04	0.02	8.65	34.09	32.83	2.52	3.16		
Off Spec Concentrate	7.60	0.36	0.18	16.55	32.92	23.05	0.92	1.49	12.33	0.10

The organisation signed an offtake agreement of sale with Glencore International, in China, for the sale of concentrates with the chemical specification shown in Table 1.2. The table shows that the level of MgO in the concentrates that are sold to Glencore International should not exceed 20%. Glencore International smelter specification allows a maximum of 5% MgO in the concentrates. However, concentrates with higher MgO content are accepted at a penalty and they are blended with other concentrates with very low MgO content before smelting.

Table 1.2 Glencore International concentrate chemical specification

Element	Specification
Ni	8% min
Cu	≈ 0.4%
Co	≈ 0.3%
Fe	≈ 20%
S	≈ 15%
Au	< 1g
Pt	< 1g
Pd	< 1g
MgO	< 20%
Al_2O_3	≈ 2%
SiO_2	≈ 30%
CaO	≈1%

Smelter penalties have been set at US$2 per dry metric ton; for every additional 1% MgO content, for all concentrates that have the MgO content that is greater than 5%. Thus, all concentrates that have MgO content that ranges from 6-20% are penalized according to the penalty agreement. The maximum target MgO content for the feed blend for the BSR was set at 12% while it was still in operation (Dzingayi, 2006). However, some blended concentrates had the MgO content of up to 17% and this resulted in higher operating temperatures that were greater than 1500^{o}C during the smelting process (Dzingayi, 2006).

Other smelters like Botswana Colliery Limited (BCL), Kalgoorie in Australia and Jinchuan in China operate with lower MgO levels in their slag, as indicated by their lower slag skimming temperatures. Their slag skimming temperatures ranges from 1244^{o}C-1380^{o}C, with the average MgO content of 7% in the slag (Warner et al., 2007). The slag skimming temperatures at BSR were in the range of 1460^{o}C-1580^{o}C due to the high MgO content in the slag (Dzingayi, 2006). Hence; there is a need to reduce the levels of MgO in the concentrate to 12% and below. This will lead to reduced penalties and also meet the BSR smelter feed specification that will enable them to operate at normal temperatures of around 1390^{o}C -1400^{o}C when the operations resume.

The mineralogical analysis of Trojan's feed and concentrates, with all the magnesia containing silicates highlighted, are shown in Table 1.3. The analysis shows that pyroxene is the major contributor of MgO in the concentrates, followed by the smectites, chlorites and serpentine.

Table 1.3 Mineralogical analysis of Trojan's feed, concentrate and tailings

Mineral	Feed Wt. %	Concentrate Wt. %	Tails Wt. %	Recovery %
Pentlandite	4.35	32.43	0.17	97
Maucherite	0.07	0.11	0.04	20
Pyrrhotite	6.93	28.86	3.99	54
Cu Sulphides	0.23	1.85	0.01	100
Other Sulphides	0	0.11	0	
Olivine	41.6	7.18	27.93	2
Pyroxene	5.97	16.74	3.65	36
Chlorites	15.12	6.13	17.93	5
Amphiboles	8.24	1.33	10.04	2
Mica	2.04	0.13	1.47	1
Smectites	0.04	0.06	0.03	19
Serpentine	3.53	1.06	20.86	4
Quartz	2.12	0.27	2.46	2
Feldspar	4.04	0.43	5.07	1
Garnet	1.77	0.68	1.91	5
Other	3.95	2.63	4.43	9
Total	100	100	100	

Most nickel plants use depressants during their flotation stages in order to reduce magnesia content that reports to the concentrate. The reagents that can be used to depress the MgO bearing silicate minerals include cellulose compounds (carboxy methyl cellulose), polysaccharides (guargum and starches), polyacrylamides, nigrosine-dextrin type compounds and tannin rich compounds. Eltham and Tilyard (1973) tested a number of these reagents on West Australian nickel sulphide ores and they concluded that carboxy-methyl cellulose (CMC) and guargum were the most effective talc depressants. For chlorides, magnesite and dolomite, guars and the dextrins showed a good depressing effect (Bulatovic, 2007). However, Rhodes (1981) showed that any further additions of guargum above the optimum dosage depressed the sulphides instead. Hence, to reduce the MgO in concentrates, flotation tests will be carried out using the CMC and guargum depressants, as well as different collector combinations.

From the daily laboratory test results, it was noted that there was overgrinding taking place in the Trojan's milling circuit especially for the feed with a nickel content greater than 2%. The MgO removal or desliming cyclones are mainly supposed to take out MgO slimes, however, the nickel content in the overflow is sometimes higher or almost equal to that of the underflow. The target grind size from the primary milling circuit is 50-55% passing 75µm, however, the grinds of up to 80% passing 75µm have been obtained in the primary milling circuit.

Overgrinding results in most of the finer particles of nickel reporting to the tailings stream, and the tailings assays, as high as 0.4% Ni have been recorded. Therefore, there is also a need to optimize the primary milling circuit in order to meet the target grind size, and to reduce nickel losses into the tailings.

1.1 Problem Identification

The presence of high MgO in the concentrates is detrimental to the upstream smelter operations. A high MgO content (>16%) necessitates operating temperatures of (>1500ºC) in order to maintain the required slag viscosity as compared to the normal operating temperatures of 1390ºC -1400ºC. According to the Trojan's smelter records, these operating temperatures can be achieved with the MgO content in the concentrate that is less than 8%. However, at higher temperatures, the attack on the frozen slag layer between the bricks can take place leading to furnace breakouts and reduced lining life. For example; on the 6th of May 2002, a break-out through the hearth of the six-in-line electric arc furnace occurred at the Trojan's smelter. The feed blend contained 16% MgO which resulted in high MgO in the slag, above 20%, which required temperatures greater than 1500ºC to flow (Dzingayi, 2006). Operating at such high temperatures results in increased energy costs and also damages the auxiliary furnace equipment which can lead to increased downtimes. Since furnace relining is one of the major costs, this can lead to a substantial increase in the smelter operating costs and increase in loss of revenue due to down times. The average MgO content in Trojan's concentrates, for the year 2015, was 16.14%, and 8000 to 9000 tonnes are shipped per month. The higher MgO content in the concentrates has led to significant losses in monthly revenue, estimated at US$141 000 due to the smelter penalties levied by Glencore International for concentrates that have MgO content greater than 5%.

1.2 Research Objectives:

The objectives of the study are

- To reduce the MgO levels in the nickel concentrate to less than 12% by coming up with the best flotation reagent suite to depress MgO.
- To optimize the primary milling circuit in order to meet the target grind size, and to reduce nickel losses into the tailings.

Key Questions

1. Which collector combinations and depressant dosages can effectively reduce the recovery of MgO, mainly pyroxene, into the Trojan's concentrates?
2. Which particle size classes results in the highest nickel recovery and lower MgO content in the concentrates?

1.3 Format of the Research Report

The best reagent suite was determined from laboratory flotation test results using the CMC and guargum reagents from different suppliers as well as different collector combinations. Reagent suites need to be determined for specific ores, since all nickel ores differ in mineralogy. The floatable gangue minerals in nickel ores differ with different ore deposits. Hence, there was a need to come up with the reagent suite that can effectively reduce the recovery of, mainly, pyroxene into the Trojan's concentrates. The MODSIM simulator was used to determine the milling parameters for disseminated ores that gave the optimum particle size distribution (PSD) for the coarse flotation circuit feed. The Plitt model equations were used predict the performance of primary cyclones, the mass balance as well as the size distributions of the products of the cyclone. It is important to ensure that the hydrocyclone classifiers are operating properly, to obtain the predetermined PSD into the flotation circuit.

1.3.1 Delimitations of the Research

This research focused on finding the best reagent suite that could effectively depress MgO in the Trojan's nickel ores as well as the optimization of the primary milling circuit.

1.4 Format of the Research Report

Chapter 2 gives an overview of literature on the geology and mineralization of Trojan Nickel Mine ores, process overview of the concentrator, flotation process and reagents, modelling and simulation of the primary milling circuit and the hydrocyclone operation. Chapter 3 gives details

on the methods and materials used in carrying out the proposed test work; flotation tests using CMC and guargum depressants, and the selection and breakage function tests.

Plant survey and MODSIM based simulation results are given in Chapter 4. Flotation test results, given in Chapter 5, consist of the depressants screening tests, flotation test without a depressant and collector combination test. The conclusion and recommendations on the results obtained are given in Chapter 6. The outline of the references used in the research are given in Chapter 7. The Appendix, in Chapter 8, gives the simulation data and results, as well as some flotation test results.

CHAPTER 2

LITERATURE REVIEW

In order to gain insight into the study, the literature below has been reviewed so as to come up with the best approach to solve the problem identified.

2.1 Introduction

The Trojan nickel deposit, 90km north of Harare, occurs in serpentinized Archaean ultramafic lavas of the Mazoe greenstone belt (Chimimba and Ncube, 1986). The sulphides are hosted in a sequence of komatiitic ultramafic lavas (Chimimba and Ncube, 1986). The main sulphide minerals in order of abundance, are pyrrhotite, pentlandite and chalcopyrite. Minor amounts of millerite and pyrite, and the arsenides, maucherite, niccolite, cobaltite and gersdorffite are present (Chimimba and Ncube, 1986). The mineralization occurs as disseminated, massive or near-massive ores. The basal position of the near-massive ore and layers of massive ore relative to the disseminated ore, and the decrease in the nickel content from the footwall of the ore bodies to the hanging wall, suggests the ore bodies are of magmatic origin and were formed by gravity settling of the sulphides during crystallization of the ultramafic lavas (Chimimba and Ncube, 1986).

The Trojan's massive ores occur as a layer of sulphides up to 1m thick along the contact between the mineralized serpentinite and the basal iron-formation or quartz-feldspar porphyry (Chimimba and Ncube, 1986). They consist of 60-90% sulphides and around 10% nickel (Hofmann et al., 2013).

The near-massive or net-textured ore occurs as sulphide-rich lenticular serpentinite pods at the base of a mineralized serpentinite flow and has a sharp contact against the overlying disseminated ore (Chimimba and Ncube, 1986). Near-massive ores consist of 30-40% sulphides and around 4% nickel (Hofmann et al., 2013).

Disseminated nickel ores (average 0.6% Ni) are pervasive and makes up ~ 95% of the reserves (Hofmann et al. 2013). The disseminated sulphide ores occur as discrete blebs up to 1cm wide in

serpentinite (Chimimba and Ncube, 1986). In talc seperntine or talc schist the sulphides frequently occur as stringers along foliation planes (Chimimba and Ncube, 1986).

The nickel ore is mined from underground where it undergoes primary crushing before being hoisted to the surface. The mineral beneficiation stages at the Trojan's concentrator include crushing, milling and flotation. The purpose of the flotation process is to separate the unwanted gangue minerals from the valuable minerals to give the highest possible recovery and highest possible grade.

Bulk flotation of sulphides is carried out to separate the valuable sulphide minerals which are pentlandite, chalcopyrite and pyrrohotite from the gangue minerals which are mainly MgO bearing silicates, quartz, feldspar and garnet. This is done by minimizing the recovery of the gangue minerals into the concentrates by the use of the flotation reagents. These gangue minerals when recovered to the concentrate, reduce the concentrate grade and the high MgO levels lead to higher operating temperatures at the smelter. Hence; the need to minimize their content in the final concentrate coming out of the flotation circuit.

The comminution stages and screening are carried out to reduce the size of the ore for further processing. These include the primary (-125mm), secondary (-19mm) and tertiary crushing (-8mm) stages which progressively reduces the ore to indicated particle sizes. The ore that is less than 8mm is placed on a stock pile before it is fed into the primary mills where the liberation of valuable minerals takes place.

2.1.1. Trojan's Milling Circuit

The Trojan primary milling circuit consists of three ball mills (1, 2 and 4) as shown in Figure 2.1. It also consists of the regrind mill number 3, shown in Figure 2.2, which is operated in closed circuit with 350mm hydrocyclones. Table 2.1; shows the technical details of ball mills used in the milling circuits.

Table 2.1 Technical details of Trojan ball mills

	Primary Circuit		Regrind Circuit
	MILL 1	**MILL 2**	**Mill 3**
Mill Diameter (m)	3.66	3.66	2.75
Mill Length (m)	4.88	4.27	3.66
Manufacturer	Bateman	Bateman	Craster
Type	Overflow	Overflow	Overflow
Mill speed (RPM)	16.1	16.3	18.2
% of Critical Speed	72.9	73.8	71.4
Liner Profile	Single wave	Single wave	Single wave
Ball Sizes (mm)	80	80	40
Grinding Media Type	Balls	Balls	Balls
Grinding Media Material	Cast steel	Cast steel	Cast steel
Type of Lining Material	Mn-Steel	Mn -Steel	Rubber

2.1.1.1 Primary Milling

Ball mill 1 and 2 operate in closed circuit with hydrocyclone classifiers while ball mill 4 operates in an open circuit with the hydrocyclones. Milling is carried out using ball mills at 72% -78% solids. The ground ore is discharged from the mill through an 18mm aperture sized trommel screen. The trommel underflow for the three mills is then fed into the primary sump. The pulp density for the primary cyclone feed is regulated at 26%-36% solids.

The mill discharge then undergoes classification using a cluster of five 600mm hydrocyclones and usually three cyclones are operated at a given time. The product is recovered as the overflow while the underflow is fed into the mills for further liberation. The overflow with particle size 50-55% passing 75µm is dewatered using the dewatering cyclones to the pulp density of 25-33% solids and then fed into the rougher flotation banks.

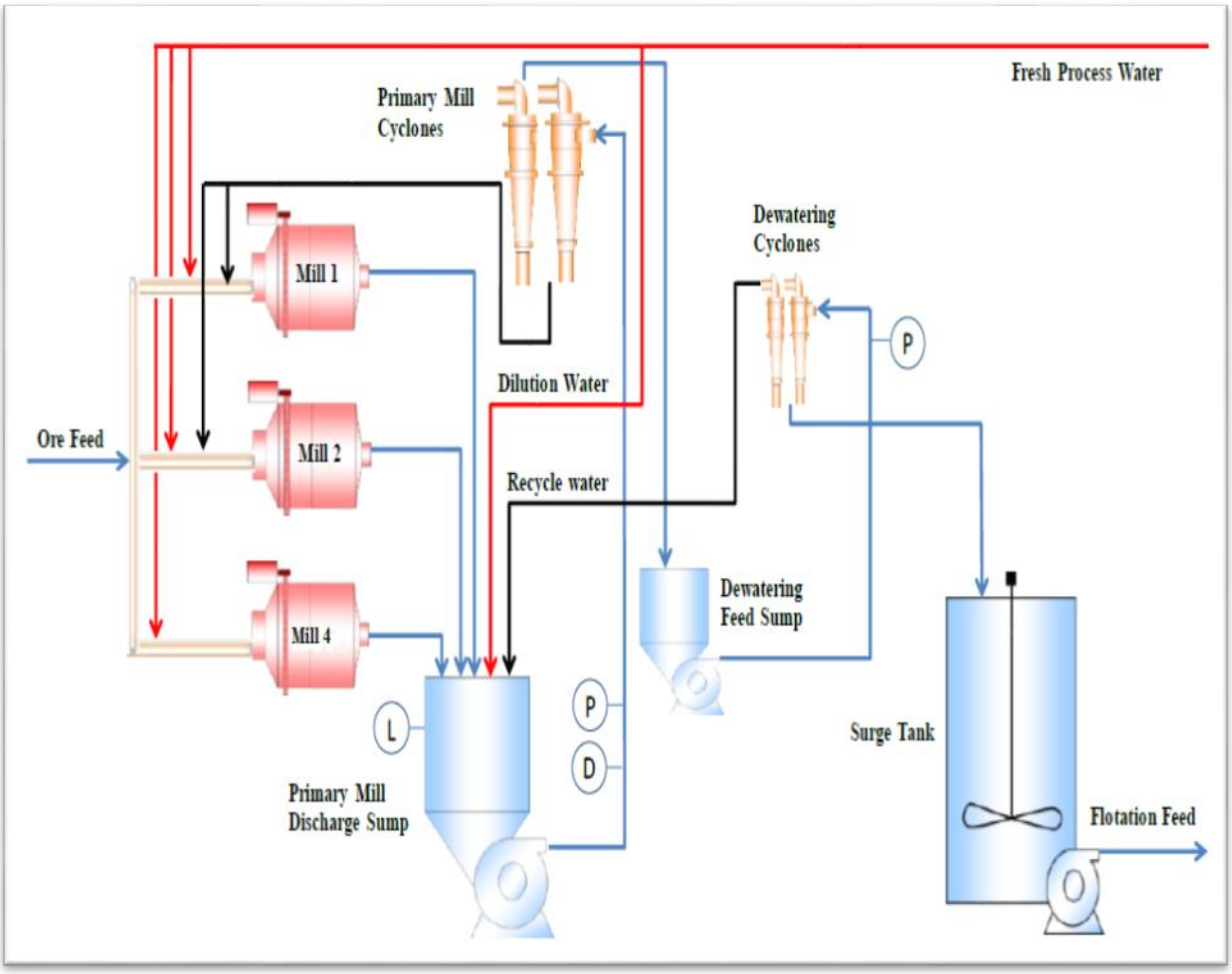

Figure 2.1 Trojan concentrator primary milling circuit

2.1.1.2 Secondary Milling

Desliming of the rougher and rougher cleaner tailings is carried out, using the MgO removal or desliming cyclones, before the secondary milling process. The cyclone underflow is then fed into the regrind mill 3, which is operated in closed circuit with the 350mm diameter regrind cyclones as shown in Figure 2.2. The target grind size for mill 3 is 60-65% passing 75µm, liberating locked, slow floating valuable minerals from the gangue. The mill product is classified and the overflow is pumped into the sump while the underflow is recirculated into the mill for further grinding.

2.1.2 Trojan's Flotation Circuit

The Trojan flotation circuit consists of cells that are connected into flotation banks. The rougher banks consist of five $30m^3$ cells. Cleaner and re-cleaner banks consist of four and three $10m^3$ cells respectively, while the final cleaner banks consist of four $5m^3$ cells. The operating parameters for each cell such as air flow rate, froth depth in each bank are different from other cells in the bank. The flotation circuit consists of a roughing stage and three cleaning stages for both the coarse and fine feed as shown in Figure 2.2.

2.1.2.1 Coarse Flotation

The underflow from the dewatering cyclones with particle sizes 55-60% passing 75µm is deposited into a surge tank where the pulp is conditioned. Pulp containing 25-33% solids is then fed into the rougher banks. A concentrate produced is taken through three cleaning stages while the tailings are taken for regrinding in the secondary milling circuit. Rougher cleaner tailings are also taken to the tailings sump while awaiting classification and regrinding. The concentrate from the cleaner banks is taken through the re-cleaning and final cleaning stages. Three cleaning stages are carried out to produce a higher concentrate grade and to reject misplaced gangue particles. Tailings from the re-cleaner bank are recycled into the rougher cleaners, while the final cleaners tailings are recycled into the re-cleaner bank. The final cleaners concentrate is collected into the final concentrate sump.

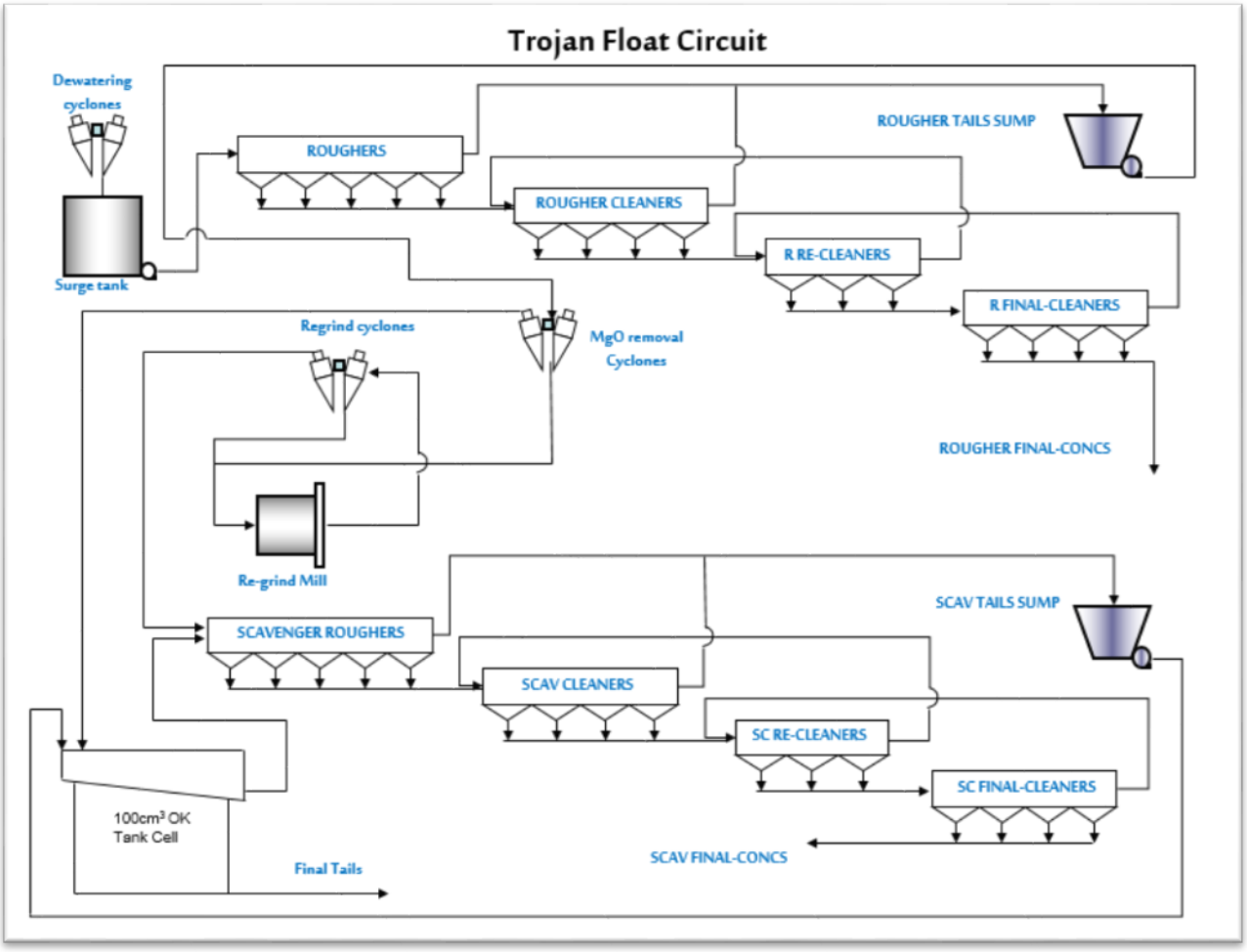

Figure 2.2 Trojan concentrator flotation and regrind circuits

2.1.2.2 Fine Flotation

The regrind cyclone overflow with 23-25% solids is fed into the scavenger rougher bank. The scavenging stage aims at improving the flotation circuit recovery and minimizing losses of valuable minerals to the tailings. Scavenger rougher concentrate is taken through cleaning stages as shown in Figure 2.2. Scavenger rougher and scavenger cleaner tails are collected into the scavenger tailings sump, and fed into the 100m^3 Outokumpu tank cell in order to recover fine valuable minerals. The Outokumpu cell concentrate is fed into the scavenger cleaner bank and the tailings are discharged to the 150ft tailings thickener. Scavenger rougher concentrates are then taken through the scavenger cleaners, re-cleaners and final cleaners stages.

The scavenger re-cleaners tailings are recycled into the scavenger cleaners, while scavenger final cleaner tails are recycled into the scavenger re-cleaners bank. The final concentrates from the rougher and the scavenger banks are then combined and sent to the 80ft thickener.

Thickened concentrates are then filtered using the drum and belt filters; where the moisture content is reduced to 15%. After the filtration stage, the concentrates are dried to 4-5% moisture content using the coal-fired rotary kiln dryer. The dried concentrates are then bagged and shipped to Glencore International in China.

2.1.2.3 Depressant Addition Points

Stage addition of the Betamin depressant is carried out in the flotation circuit. Table 2.1; shows the number of depressant dossing points in each flotation bank. These points are the same for all the rougher and scavenger banks. The dosages are varied according to the operator's discretion based on the visual judgment of the head grade.

Table 2.2 The number of depressant dossing points in the flotation circuit

Flotation Bank	Dosing points (cell)	Number of Cells	Cell Capacity m^3
Rougher banks	1 and 3	5	30
Cleaner banks	1	4	10
Re-cleaner banks	1	3	10
Final cleaner banks	1	4	5
Outokumpu cell	1	1	100

2.2 Froth Flotation

Froth flotation is a concentration method that is used for selective separation of mineral species from the slurry that consists of both valuable and gangue minerals (Gorain et al., 2000). Separation is effected by the difference between induced surface conditions on the valuable minerals and gangue minerals (Wills and Napier-Munn, 2007). Conditioning of the slurry using different flotation reagents is carried out prior to flotation to impart differential hydrophobicity between the valuable and gangue minerals (Wills and Napier-Munn, 2007). The valuable minerals are rendered hydrophophic by use of a collector so that they can get attached to bubbles in the cell.

The slurry is agitated using the impellor to keep all the solids in suspension and disperses the air introduced into the cell into fine bubbles (Gorain et al., 2000). Particles that get attached to air bubbles are carried to the surface, collected in the weir, while the completely wetted particles remain in the slurry. Froth flotation has been found to be a useful treatment method for fine-grained ores that cannot be processed by gravity separation (Wills and Napier-Munn, 2007).

The activity of mineral surface in relation to flotation reagents in water depends on the forces that operate on that surface (Wills and Napier-Munn, 2007). Equilibrium is established between the solid-air, solid-liquid and liquid-air interfacial tension when all the three phases are in contact (Gupta and Yan, 2006). These tensile forces lead to the development of the contact angle between the mineral and the bubble surface.

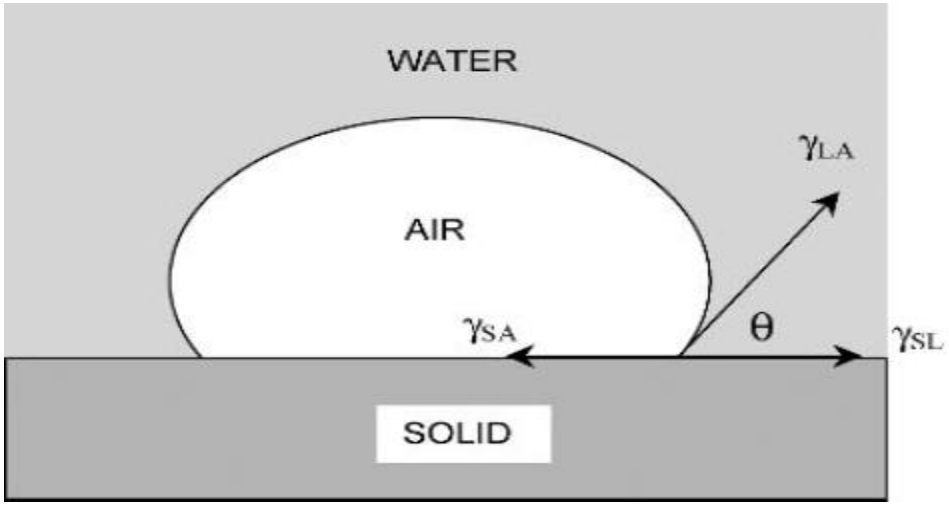

Figure 2.3 Three phase contact between solid, liquid and air in a flotation system (Gupta and Yan 2006)

The Young's equation below gives the balance of surface forces at equilibrium as shown in Figure 2.3.

$$\gamma_{SA} = \gamma_{SL} + \gamma_{LA} \cos\theta \quad \ldots\ldots (1)$$

Where:

γ_{SA} = surface energies between solid and air

γ_{SL} = surface energies between solid and liquid

γ_{LA} = surface energies between liquid and air

θ = contact angle between the mineral surface and bubble surface

The contact angle is the measure of how well the air bubble spreads on the mineral surface. A low contact angle that is less than 90^o indicates that the surface is less hydrophobic while angles greater than 90^o indicate that the surface is more hydrophobic (Gupta and Yan, 2006). A hydrophobic surface favours contact with air than water due to their lower free energy and they readily stick to available air bubbles. Minerals with greater contact angles have higher floatability and this results in higher flotation rate constants. In practice, the bubble attachments under agitation and aeration are dynamic processes that cannot be analyzed based on the equilibrium contact angle (Ralston and Neuxcombe, 1992). Leya and Poling (1960) concluded that the contact angle is simply an indication of the extent to which the given solid-liquid-air system utilizes the free energy of interfaces in the bubble deformation system.

The forces required to break the particle-bubble interface is called the work of adhesion, $W_{SA.}$ It is equal to the work required to separate the solid-air interface and produce a separate air-liquid and solid-liquid interfaces (Wills and Napier-Munn, 2007).

$$W_{SA} = \gamma_{WA} + \gamma_{SW} - \gamma_{SA} \ldots\ldots (2)$$

Combining the equations (1) and (2) gives:

$$W_{SA} = \gamma_{WA}(1 - \cos\theta) \ldots\ldots (3)$$

An increase in the contact angle increases the work of adhesion, hence the forces that hold the bubble and solid together will be greater (Gupta and Yan, 2006).

2.2.1 The Kinetic Model of Flotation

The kinetic model for flotation describes the batch flotation process in a well stirred flotation environment where the solid particles are kept in suspension. The rate of flotation is a measure of change in concentration of floatable material in the pulp at a given time.

Different particles have different specific rates due to the variation in the contact angles of the particle surfaces (King, 2001).

The batch test is described by:

$$\frac{dC_{ij}}{dt} = -K_{ij}C_{ij} \ldots\ldots(4)$$

Integrates to

$$\frac{C_{ij}(t)}{C_{ij}(0)} = \exp(-K_{ij}t) = 1 - R_{ij} \ldots\ldots(5)$$

Where:

C_{ij} (0) = concentration of floatable particles of type ij in the cell at time t = 0

C_{ij} (t) = concentration of floatable particles of type ij in the cell at time t.

t = flotation time

K_{ij} = flotation rate constant of the particles of type ij.

ij = refer to particle size and particle composition respectively.

R_{ij} = recovery of floatable particles of type ij in the cell

Non-floating minerals have $K_{ij} = 0$

Plotting ln (1-R_{ij}) against time will give K_{ij} as the slope of the straight line obtained.

2.2.2 Cell Operation

A mechanical flotation cell consists of a tank fitted with an impeller or rotor as shown in Figure 2.4. The impeller agitates the slurry to keep the mineral particles in suspension. It also disperses air into fine bubbles and provides an environment for interaction of bubbles with hydrophobic minerals (Gorain et al., 2000). The attached bubble-particle aggregates are buoyed upwards and are removed from the cell lip into the concentrate launder.

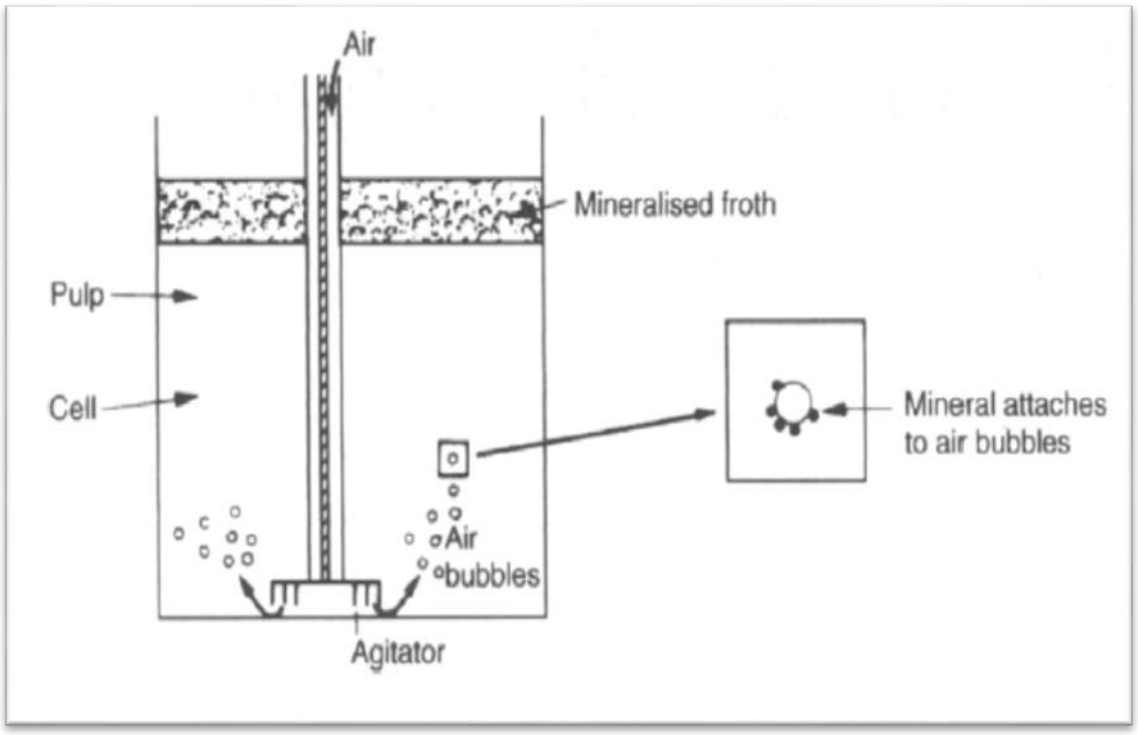

Figure 2.4 Mechanical flotation cell that is agitated from the top (Wills & Napier-Munn 2007).

Three hydrodynamic zones must be generated in the cell for effective flotation to take place (Gorain et al., 2000):

1. **Turbulent Zone**: This is the region close to the impeller, where the turbulence is required to keep the solids in suspension. The gas is also dispersed into bubbles and bubble-particle interaction takes place to enhance the collection of hydrophobic minerals.
2. **Quiescent Zone**: It lies above the turbulent zone, where bubble-particle aggregates rise up in a less turbulent region. Upgrading of the concentrate takes place as some of the mechanically entrained gangue minerals between the bubbles drains off.
3. **Froth Zone**: The zone lies above the quiescent zone. It serves as an additional cleaning stage through froth drainage and improves the grade of the concentrate.

2.3 Factors Affecting the Rate of Flotation

The flotation rate constant can be used to quantify the effect of different variables on the flotation process. Some of the flotation variables are impeller speed, air flow rate, particle size and pulp density (Gupta and Yan, 2006).

2.3.1 Impeller Speed

At a given air flow rate in a cell, an increase in impeller speed leads to more dissolved air in solution. This results in increased bubble precipitation which provides a favourable preliminary step to collision and adhesion of mineral particles to the bubbles (Gupta and Yan, 2006). Undissolved air is then broken into finer bubbles. The rate constants of all mineral particles are increased to a point where agitation is so intense that it disrupts bubble-particle attachments. Maximum rate is achieved at 1300 rpm before it starts to decline in top agitated mechanical cells (Gupta and Yan, 2006).

2.3.2 Air Flow Rate

For a given impeller speed, the increase in the air flow rate increases the rate constants of all species. The rate increases up to a particular flow rate before it starts to decrease. When more air is forced through the impeller, its residence time in the shear zone is reduced. This causes larger bubbles to form and the value of the rate constants may remain unaffected (Gupta and Yan, 2006).

2.3.3 Particle Size

The probability of collision and adhesion of a particle varies with particle size because of (Gupta and Yan, 2006):

- Its projected area
- Its inertia which determine the ability of the particle to cut across flow lines around the bubbles
- The possibility of detachment after the attachment of the particle due to disruptive turbulence
- The extent to which the collision can distort the bubble and alter the time of contact

The coarser and heavier particles may be concentrated at the lower part of the cell and their chances of collision with air bubbles are greatly reduced. The maximum floatable sulphide ore particle in mechanical cells is 420µm (Gupta and Yan, 2006). Once the mineral particle gets attached, its probability of being detached decreases with a decreasing particle size. In the

flotation of sulphides the maximum rate is realized at intermediate particle sizes of around 35 µm (Gupta and Yan, 2006).

2.3.4 Pulp Density

A variation in pulp density can affect flotation results significantly. High pulp densities can inhibit the dispersion of air and good bubble formation. This in turn affects the recovery from the flotation system. Lower pulp densities tend to produce higher concentrate grades by promoting better froth drainage. If there are large fluctuations in pulp density, the concentration of reagents added at a given time will vary greatly and this can affect the recoveries in the system (Gupta and Yan, 2006).

2.4 Recovery Mechanisms of MgO into the Concentrate

Studies by Wellham at al., (1992) showed that magnesium oxide can be recovered into the concentrate through the following mechanisms:

- **Naturally floating**: these are fast floating particles that can be recovered into the nickel concentrate without adding a collector due to natural hydrophobicity. These particles have a high contact angle and they readily get attached to the air bubbles and are recovered to the concentrate by true flotation (Witney and Yan, 1996).
- **Slow floating due to activation**: these are fairly slow floating gangue minerals that can be recovered after being activated by the addition of a collector in the flotation cell. When the collector coats these particles they are rendered hydrophobic and they get recovered into the concentrate by true flotation (Witney and Yan, 1996).
- **Slime coating on sulphide particles**: these gangue particle minerals are recovered when they get attached to floating sulphide minerals (Witney and Yan, 1996). This decreases the concentrate grade.
- The extent to which the slimes coatings occur is controlled by the magnitude and sign of the surface charge of both the particle and slime. These are reflected by the zeta potential measurements. Slimes coatings can be removed by dispersants which adsorb preferentially onto slimes, reversing their surface charge and preventing electrostatic attraction (Pietrobon et al., 1997).

- **Composite particles**: these are recovered through true flotation with sulphide minerals (Witney and Yan, 1996). This usually takes place with particles that are not fully liberated. The recovery of course sized MgO mineral particles (i.e. ~ 40µm) may be due to composite flotation, as its rate constants are similar to those of equivalent nickel size fraction (Pietrobon et al., 1997).
- **Entrainment**: this is a process whereby very fine gangue minerals that follow the water stream ends up in the concentrate or get recovered into the concentrate due to poor froth drainage (Pietrobon et al., 1997). Entrainment is more predominant for particles that are less than 10µm. This process is not selective to mineral surface properties, both gangue and valuable minerals are recovered alike into the concentrate (Wills and Napier-Munn, 2007). Drainage of entrained particles takes place in the froth phase hence the need for a sufficiently stable froth that can allow drainage of the entrained particles.

The mineralogical analysis carried out by Betachem (Grobler, 2014) on Trojan's feed, concentrate and tailings, showed that pyroxene was the major contributor of MgO into the final concentrates. Pyroxene had the recovery of 36% into the final concentrate followed by chlorite with a recovery of 5% and olivine with the recovery of 2% into the concentrate. Pyroxene contributed 50% of the MgO content in the final concentrate followed by 31% olivines and 12% chlorites. The 81% of pyroxene in the concentrate was fully liberated and within the floatable size range of 10 to 150µm. It was concluded that pyroxene was recovered by true flotation into the Trojan's nickel concentrates.

2.5 Flotation Reagents

Flotation reagents are the most important aspects of the flotation process. They make the separation of valuable minerals from gangue minerals possible by imparting the desired surface properties to the minerals (Wills and Napier-Munn, 2007). Chemical modification on mineral surfaces is achieved by addition of a number of flotation reagents. These reagents, grouped according to their functions are; collectors, frothers, regulators and depressants (Bulatovic,

2000). The best reagent suite gives the most effective separation and concentration results in the flotation process.

The selection of the reagents is largely dependent on the specific mineral mixtures being treated. These reagents are needed to control the relative hydrophobicity of the particles, and to maintain the proper froth characteristics.

2.5.1 Collectors

Collectors are organic chemicals that make the surfaces of selected valuable minerals hydrophobic. They selectively form a hydrophobic layer on certain mineral surfaces in the pulp. This provides suitable conditions for attachment of hydrophobic particles to the bubbles to take place on contact causing the particles to be recovered during the flotation process (Gupta and Yan, 2006). Collectors are added to the pulp followed by conditioning time under agitation to allow for the collector adsorption onto the mineral particles. Laboratory tests have shown that the flotation rate constant increases with increase in collector concentration up to a maximum. In continuous plants, the longer the pulp remains in the circuit, the larger the contact angle becomes, resulting in an increase in the rate constant (Gupta and Yan, 2006).

The principal collectors used in the bulk flotation of nickel sulphide ores are xanthates and to a lesser degree dithiophosphates and mercaptans (Bulatovic, 2007). Studies have shown that mercaptans and dithiophosphates are highly effective collectors for pentlandite and also highly selective towards pyrrhotite. Experimental results of Mt Keith ore indicated that the use of mercaptans with lower-chain xanthate gave a better recovery than xanthate alone (Bulatovic, 2000). Mercaptan collectors showed good selectivity and good nickel recovery on Mt Windarra ore (Bulatovic, 2000). Xanthate alone gave high nickel recovery and a low concentrate grade.

The best metallurgical results were obtained when using a combination of sodium ethyl xanthate and mercaptan R407 at a ratio of 1:1. In the presence of naturally floatable gangue, dithiophosphates or mercaptans are better collectors than xanthates especially in the presence of CMC depressants (Bulatovic, 2000). Different collector combinations and dosages will be tried on the Trojan ore in order to come up with a more selective collector combination.

2.5.2 Frothers

Frothers are organic chemicals which reduce the surface tension of water in order to stabilize the bubbles in the froth layer (Wills and Napier-Munn, 2007). They are heteropolar surfactants such as alcohol or polyglycol ethers. The frothers adsorb at the air-water interface and this results in the reduction of the water surface tension (Gupta and Yan, 2006). Smaller bubbles are then produced and these stabilize the froth when it reaches the top of the slurry until the concentrate is removed from the cell. Increasing the frother concentration within operating limits will increase the rate constants of all particles (Gupta and Yan, 2006). At high concentrations, adhesion of bubbles to hydrophobic minerals is reduced and this leads to reduced recoveries. A reasonable stable froth that allows selective drainage of entrained gangue minerals should be formed to increase flotation kinetics.

A good frother should have no collecting properties and produce a stable froth to facilitate transfer of concentrate from the cell to the launder (Wills and Napier-Munn, 2007). The choice of frother in bulk flotation of nickel sulphide ores depends on the nature of the ore as well as the gangue slimes present. In the presence of clay slimes, cyclic alcohol (pine oil) mixed with glycol-type frother gives good metallurgical results (Bulatovic, 2000). Glycol-type frothers are selected where frothing problems caused by high viscosity exist (Bulatovic, 2000).

2.5.3 Regulators or Modifiers

Regulators are organic or inorganic chemicals used to modify the slurry conditions to enhance the difference in surface chemistry between valuable and gangue minerals (Gupta and Yan, 2006). They consist of depressants, activators, pH regulators and dispersants. Modifiers are mainly used in the differential flotation of a mixed ore, where successive removal of two or more valuable minerals can be achieved by flotation (Gupta and Yan, 2006).

2.5.3.1 Depressants

Depressants are chemicals which inhibit the adsorption of a collector onto a particular mineral surface thus preventing its flotation.

They have the opposite effect of activators, and they are used to increase selectivity by preventing one mineral from floating, while allowing another mineral to float unimpeded (Bulatovic, 2007). Thus, a depressant improves the rate constant differential between the valuable minerals and the naturally floatable gangue minerals (Bulatovic, 2007).

The types of depressants that can be used to depress the MgO bearing silicate minerals are cellulose compounds (carboxy methyl cellulose), polysaccharides (guargum and starches), polyacrylamides, nigrosine-dextrin type compounds and tannin rich compounds. Polymeric depressants have an advantage of being less hazardous as compared to inorganic depressants. Carboxy methyl cellulose (CMC) and guargum are normally used to depress naturally hydrophobic minerals that are found in ultramafic sulphide ores. The major difference between the two is that CMC is negatively charged in solution while guar is slightly negatively charged (Mackenzie, 1986). Rhodes (1981) postulated that hydrogen bonding was the adsorption mechanism of CMC and guargum on readily floatable silicates. He believed that hydrogen bonding takes place between unsubstituted hydroxyl groups on the depressant molecules and the surface oxygen sites at the broken edge structures of the layer silicates. These anionic groups then render the surfaces of the silicate minerals hydrophilic, thus depressing the minerals (Mular et al., 2002).

Test results obtained on the Windarra (Australia) ore showed that a high molecular-weight CMC with the lowest sodium glycolite content gave the best talc depression and the highest concentrate grade (Bulatovic, 2007). The high molecular-weight CMC also had a depressing effect on the pyrrhotite present in the ore. The depressing effect of CMC was dependent on the pH used in flotation. In these tests, SO_2 and soda ash were used as pH modifiers and the obtained data indicated that at a higher pH (i.e. 9-10), the depressing effect of CMC improved considerably (Bulatovic, 2007).

Addition of soda ash and CMC was found to improve the flotation rate and recovery of pentlandite (Pietrobon et al., 1997). CMC group of compounds are high molecular weight anionic polymers, with a number of hydroxyl and carboxyl groups as side chains.

The presence of these groups ensures a stable adhesion of the polymer onto the MgO minerals by hydrogen bonding (Pietrobon et al., 1997).

CMC may act as a stabilizer by preventing Van der Waals forces of attraction between particles which would otherwise cause the system to coagulate in the absence of CMC (Pietrobon et al., 1997).

Recent studies have shown that the depressing effect of carboxy methyl cellulose can be greatly improved in the presence of aluminum chloride. It is postulated that improved depression of talc is due to the formation of $Al(OH)_3$ on the mineral surfaces (Bulatovic, 2007). Mani (1997) illustrated the negative influence of serpentine on pentlandite recovery, by adding 30% by weight serpentine to the pentlandite ore in a laboratory mill. This resulted in reduced nickel recovery, from above 80% to values below 30%. From the literature above it can be noted that the presence of these gangue silicate minerals can reduce the recovery of the valuable minerals. The CMC and guargum depressants will be tested on Trojan ore and the effect of raising the slurry pH on nickel recoveries will also be investigated.

2.5.3.2 Activators

An activator alters the chemical nature of the mineral surfaces in order to enhance the adsorption of a collector (Gupta and Yan, 2006). Activators are generally soluble salts which ionize in solution and their ions then react with the mineral surface to form a thin film. These reagents make it possible for collectors to adsorb onto surfaces that they could not normally attach to. They enhance the adsorption of the collector on the selected minerals. For example, sphalerite is activated by the use of copper sulphate. The collector used in the flotation of sphalerite, such as zinc xanthate are relatively soluble in water and do not form a hydrophobic film around the mineral. Activation is achieved by the formation of copper sulphide on sphalerite since copper is more electronegative than zinc.

The copper sulphide deposited on sphalerite reacts with xanthate to form an insoluble copper xanthate (Wills and Napier-Munn, 2007). This renders sphalerite hydrophobic. The activator used in the bulk flotation of sulphides is copper sulphate.

This can lead to activation of all sulphides including pyrrhotite, and, this can reduce the concentrate grade significantly. The cost of rejecting pyrrhotite will also be weighed against the shipping costs of the concentrate.
This would be more beneficial when most of the pyrrhotite in the ore contains insignificant amounts of nickel. The overall recovery of nickel in the plant will not be adversely affected when pyrrhotite with very low nickel content is rejected.

2.5.3.3 pH Regulators

These reagents are used to adjust the pH of the pulp to give an optimum performance for a particular reagent and mineral (Gupta and Yan, 2006). The surface chemistry of most minerals is affected by the pH. In general, minerals develop a positive surface charge under acidic conditions and a negative charge under alkaline conditions. Since each mineral changes from negatively-charged to positively-charged at some particular pH, it is possible to manipulate the attraction of collectors to their surfaces by pH adjustment.

The typical regulators used are lime, soda ash and sulphuric acid. Lime and soda ash are often added to the slurry prior to flotation to precipitate the heavy metal ions from solution. When xanthate collectors are used, a critical pH is reached above which the valuable minerals will not float (Wills and Napier Munn, 2007). At the Thompson mill in Canada, when soda ash was used for pH control at around 9, the rate of pentlandite flotation increased relative to that of pyrrhotite as compared to the rates at a natural pH of 8 (Mular et al., 2002). Rejection of large amounts of pyrrhotite can lead to reduced shipping costs and also reduce sulphur emissions from the smelter. However, this may also lead to reduced nickel recoveries, since pyrrhotite contains fine (1-10μm) flame-like intergrowths of pentlandite in its grains (Mular et al., 2002)

In the case of Birchtree ore in Canada, soda ash was used for the rejection of pyrrhotite (Mular et al., 2002). The pH was maintained at the value of 10 in the rougher scavenger circuit and 85% of pyrrhotite was rejected with 86% recovery of pentlandite.

Eltham and Tilyard (1973) noted that the addition of soda ash increased nickel recovery, even in the absence of a collector. It also increased nickel flotation selectivity against MgO mineral recovery, suggesting that it acts as a dispersant.

Eltham and Tilyard (1973) suggested that due to addition of soda ash, the dissolved or adsorbed metal ions were precipitated as insoluble carbonate salts which would have otherwise depressed pentlandite flotation. The carbonate ions from soda ash do enhance dispersion in the pulp and the CMC may assist in the removal of adhering slime particles from pentlandite surfaces (Pietrobon et al., 1997). The extent to which slime coatings occur is controlled by the magnitude and sign of the surface charge of both the particle and slime. This is reflected by the zeta-potential measurements. Parsonage (1985) suggested that slime coatings occur when there is no potential energy barrier for heterocoagulation.

Edwards et al., (1980) investigated the effect of chrysotile and lizardite slimes on pentlandite flotation and deduced that the formation of slime coatings was directly related to the surface charge of both pentlandite and the gangue minerals. Slime coating were removed by the addition of dispersants which adsorbed preferentially to slimes, reversing their surface charge, thus preventing the electrostatic attraction (Pietrobon et al., 1997).

No pH regulator is currently used at the Trojan's concentrator. Most plants use soda ash to reject pyrrhotite especially in countries where there are stringent environmental laws associated with sulphur dioxide emissions at the smelters. Soda ash will be used in the laboratory tests as a pH regulator with the CMC depressants.

2.5.3.4 Dispersants

Dispersants are used to remove clay slimes from mineral surfaces and to prevent the fine particles from aggregating. They improve the floatability of valuable minerals by preventing slime coatings on the mineral particles (Bulatovic, 2007). Some of these reagents are used as depressants as well as activators. Slimes can either coat valuable minerals and depress them or activate gangue minerals in the flotation system. This affects the grade of the concentrate as well as the bubble surface tension in the system. The examples of inorganic dispersants are sodium silicate and sodium polyphosphates.

Organic dispersants that can be used are starches, dextrins, guars and lignin sulphonates. These are used as depressants in the same form and molecular structure as they are used as dispersants (Bulatovic, 2007).

2.6 Pre-flotation of Hydrophobic Silicate Minerals

Pre-flotation of naturally hydrophobic silicate minerals prior to sulphide flotation was undertaken by Witney and Yan in 1996. This process resulted in successful removal of talc, however, there was a significant loss of sulphide minerals to the talc concentrate. About 50% of floatable MgO was recovered in the collectorless prefloat and the other 50% was recovered in the subsequent sulphide float. The loss of sulphide minerals into the talc concentrate was caused by the activation of sulphide minerals due to the naturally hydrophobic slime coatings. Microscopic examination of the talc prefloat concentrate showed the presence of liberated sulphide particles (Witney and Yan, 1996). This was attributed to either entrainment or slimes coating of nickel particles by the naturally hydrophobic magnesia slimes.

Preflotation will not be considered as one of the possible solutions since it results in higher nickel losses into the talc prefloat concentrate. It will also be difficult to include a new processing stage in an existing concentrator plant.

2.7 Modelling and Simulation

Simulation is the process of designing a computerized model of a system, for the purpose of understanding its behavior and developing strategies to control the operation. Simulation is now an effective tool for mineral processing plants. Some of the simulation software used for designing and optimization of mineral processing circuits include: JKSimMet, MODSIM, METSIM and USIM PAC. Modular Simulator for Mineral Processing Plants (MODSIM) is one of the most basic and easy to use simulator. It calculates a detailed mass balance for any ore dressing plant. The mass balance will include total flow rates of water and solids, the particle PSD of the solid phase, the distribution of particle composition and average assays of the solid phase (MODSIM, 2004).

Simulation on the Trojan concentrator primary milling circuit will be carried out in order to determine the parameters that meet the target grind size required for the flotation circuit feed. This will eliminate overgrinding and reduce the loss of the valuable minerals into the tailings as fines. This can also result in reduced power consumption in the milling circuit. The operation of the primary hydrocyclones will also be optimized in order to get the optimum PSD from the milling circuit.

2.7.1 Selection and Breakage Function Model by Austin

Grinding is the last stage in the comminution process and is carried out in tumbling mills, where the particles are reduced in size by a combination of impact and abrasive forces. Most ball mills operate in a closed circuit with a classifier in order to minimize the extent of overgrinding and also to reduce oversized particles in the product stream (King, 2001). Grinding is treated as a rate process, breakage of the given size fraction usually follows a first-order law when efficient breakage occurs in the mill. Selective breakage occurs within a given size range. The proportion of particles in a size class that is broken is represented by S. Thus S_1, S_2, S_3…..S_N is the size fraction that would be selected for size reduction while the rest of the particles pass through without any change in size (Gupta and Yan, 2006).

In a batch grinding process, the breakage rate of material from the initial size interval $(i = 1)$ can be expressed as (King, 2001):

$$\frac{d(Wp_1)}{dt} = -S_1 Wp_1 \quad \ldots\ldots (6)$$

Rate of grinding is proportional to the mass of particles in that size class

S_1 = selection function or rate of breakage of particles of class1

P_1 = weight fraction of particles in class 1

W = total weight of particles

The first order grinding hypothesis states that the rate of grinding is directly proportional to the particles in class 1.

Assuming that S_1 does not change with time, then the equation integrates to:

$$\frac{p_1(t)}{p_1(0)} = \exp(S_1 t) \ldots\ldots (7)$$

Plotting $\ln(p_1(t)/p_1(0))$ against time, gives the rate of milling S_1, for the class 1 particles as the gradient of the straight line.

The specific rate of breakage function has a general mathematical formula proposed by Austin et al. (1984) as shown in equation (8):

$$S_i = \frac{a x_i^{\alpha}}{1 + \left(x_i/\mu\right)^{\wedge}} \ldots\ldots (8)$$

Where x_i is the upper limit of the size interval i in mm, and a, α, μ and Λ are the model parameters that depend on the properties of the material and grinding conditions. This equation allows interpolation and extrapolation to obtain estimates of S values for all size intervals involved (metso.com, Accessed 8\10\2014).

The weight fraction of the material broken from the size interval j which appears in the size interval i before re-breakage of the fragments occurs, is defined as the primary breakage distribution function, b_{ij} (metso.com, Accessed 8\10\2014).
This can be written in cumulative form as:

$$B_{ij} = \sum_{k=n}^{i} b_{kj} \ldots\ldots (9)$$

Where:
b_{ij} = mass fraction of material broken from size class j into size i.
B_{ij} = is the cumulative mass fraction of particles passing the top size interval i from breakage of particles of size j.

B_{ij} can be fitted to an empirical function (Austin at al., 1984);

$$B_{ij}(X_iX_{j)} = \phi_j\left(\frac{X_i}{X_{j+1}}\right)^{\gamma} + (1-\phi_j)\left(\frac{X_i}{X_{j+1}}\right)^{\beta} \ldots\ldots(10)$$

Where ϕ, γ and β are the model parameters that depend on the properties of the material.
Thus; B functions are not affected by different ball filling ratios and mill diameters. Determining the breakage parameters of the ore will give an insight on what takes place in the mills.
The expected size distribution in each particle class of the ore that results from the breakage can also be determined.

2.7.2 Scale up of the Austin Model Selection Function

The parameters α and Λ in the Austin model for specific rate of breakage are usually assumed to be material specific only while S_1 and μ depend on the geometrical scale. The dominant variables are the mill diameter D_m and the size of the balls that make up the media d_b. These determine the average impact energy in the mill and have a significant influence on the value of the constant S_1.
The scale up law is given by (King, 2001):

$$S_i(d) = a_T\left(\frac{X_i}{X_0}\right)^{\alpha}\left(\frac{1}{1+\left(\frac{X_i}{C_1\mu_T}\right)^{\wedge}}\right)C_2C_3C_4C_5 \ldots\ldots(11)$$

Where C_1 to C_5 are scale up factors.
Smaller ball sizes produce less energetic impacts and each impact influences fewer particles in their immediate vicinity of the impact point between any two balls, hence, they have a lesser influence on the specific rate of breakage (metso.com, Accessed 8\10\2014).

2.8 Factors that affect the Mill Product PSD

Ball mills can operate over a wide range of conditions and geometries. Most ball mills usually operate in closed circuit with a classifier and the circuit performance is determined by the interaction between these two units (Napier-Munn, 1996).

A number of factors can affect the particle size distribution of the mill discharge such as slurry density, ball size, feed size PSD, feed rate etc.

2.8.1 Slurry Density

In a well-mixed ball mill, the slurry density inside the mill will be similar to the mill discharge density. Adding more water to the mill reduces the number of smaller particles available to interact with the balls.

This reduces the number of effective impacts causing reduction in the grinding rate (Napier-Munn, 1996). Further reduction in slurry density causes the slurry to be insufficiently viscous to coat the balls and grinding will decrease rapidly. An increase in slurry density increases the grinding rate until the slurry viscosity begins to increase rapidly. This will lead to reduced impacts between the balls, hence the reduction in the grinding rate (Napier-Munn, 1996).

2.8.2 Ball Size Selection

The coarse feed particles require bigger balls to reduce them, while the finer target grind size will also require smaller ball sizes. If the ball size becomes much smaller, the liner wave height should also be reduced (Napier-Munn, 1996). In some cases two or more ball sizes are used in order to obtain the required particle size distribution.

2.8.3 Feed Size

Narayan et al., (1987) showed that a reduction in the top size particles in the feed to a large mill produced a substantial increase in throughput. An increase in the amount of top size feed particles will result in increased number of coarser particles in the mill discharge. The coarse sized particles in the mill discharge cause the most wear in the plant equipment. This problem is worsened by operating at high circulating loads (<2.5 times the new feed rate). The high flow rate into the mill will carry coarse particles into the classifier (Napier-Munn, 1996).

2.8.4 Classifier Operation

The closed circuit operation should produce fewer coarse particles without producing excessive fine product. A lower circulating load will allow a longer mill residence time and a finer ball mill

discharge. This will also allow readily finished products more opportunities to be broken further (Napier-Munn, 1996). If the circulating load is too high, it will tend to carry coarse particles out of the mill and they can cause wear problems in the cyclones.

An optimum circulating load provides a good compromise between overgrinding and minimization of coarse particles in the mill discharge. This requires matching the classifier cut-size to the mill product size with maximum water split to the overflow (Napier-Munn, 1996).

2.8.5 Feed Rate

For the maximum rate of breakage to be achieved, the fractional void filling (U) in the mill should be equal to 1. Higher feed rates into the mill will result in a coarser mill discharge product. In this case U>1, the load in the mill is expanded and the ball to ball contact is greatly reduced resulting in the reduced rate of milling. If the feed rate is too low a finer mill product will be produced (Napier-Munn, 1996). In this case U<1, the grinding zones between the balls are poorly utilized leading to finer mill discharge.

2.9 Hydrocyclone Operation

A hydrocyclone is a continuous classifying device that utilises centrifugal forces to accelerate settling rates of particles (Wills and Napier-Munn, 2007). Particles are separated according to size, shape and specific gravity. It consists of a conically shaped vessel that is open at its apex or underflow as shown in Figure 2.5.

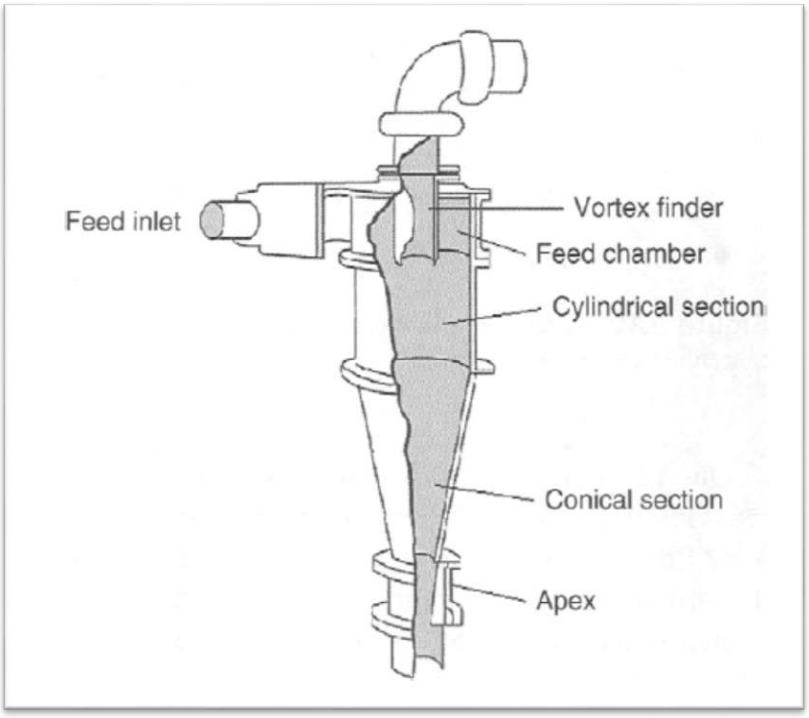

Figure 2.5 Hydrocyclone (Wills and Napier-Munn, 2007)

The conical section is joined to the cylindrical section which has a tangential feed inlet. The overflow pipe extends into the body of the cyclone by a short removable vortex finder, which prevents the short-circuiting of feed directly into the overflow (Wills and Napier-Munn, 2007). Slurry is fed tangentially into a cyclone at a given inlet pressure and the centrifugal forces developed accelerates the settling rates of particles (Gupta and Yan, 2006).

Faster settling particles move towards the wall where the velocity is lowest and they move to the apex opening. Slower settling particles are acted upon by drag forces and they move towards the low pressure zone along the axis and are carried upward through the vortex finder to the overflow (Wills and Napier-Munn, 2007). Particles thrown outside the envelope of zero vertical velocity by the greater centrifugal force exit via the underflow. Particles lying on the envelope of zero velocity are acted upon by equal centrifugal and drag forces. These particles have equal chances of either reporting to the underflow or overflow (Wills and Napier-Munn, 2007). The classification is not 100% efficient; some lighter particles get entrapped in the heavier particle stream and are lost through the apex, while some heavier particles also get lost to the overflow stream (Gupta and Yan, 2006).

2.9.1 Plitt Model for Hydrocyclones

The Plitt model for hydrocyclones is a universally applicable mathematical model that was developed after carrying out a series of experiments. This empirical model is in agreement with fundamentally derived models and it has a broad database which includes all the major variables (Plitt, 1976). The model can be used to predict the performance of cyclones with reasonable accuracy over a wide range of operating conditions without experimental data. The following design and operating variables were studied (Plitt, 1976):

- Cyclone diameter
- Height
- Inlet diameter
- Vortex finder diameter
- Apex diameters
- Feed pressure
- Solids content of the slurry feed

Four fundamental parameters must be determined as the operating and design variables, namely: Cut size d_{50c}, volumetric split between overflow and underflow, sharpness of classification m and pressure drop. When these parameters are calculated for a given set of conditions, the complete mass balance and size distribution of the cyclone can be determined.

The model consists of four basic equations:

Cut size d_{50c}: the particle size that has the 50% probability of reporting to the underflow or overflow stream.

$$d_{50c} = 50 \times \frac{5{D_c}^{0.46}{D_i}^{0.6}{D_0}^{1.21}\exp(0.063\phi)}{{D_u}^{0.71}h^{0.38}Q^{0.45}(\rho_s - \rho)^{0.5}} \ldots\ldots (12)$$

Volumetric split between overflow and underflow: the volumetric split between the overflow and underflow is used to determine the water balance across the classifier.

$$S = 1.9\frac{\left(\frac{D_u}{D_0}\right)^{3.31} h^{0.54}({D_u}^2 + {D_0}^2)^{0.36} \exp[0.0054\phi]}{{D_c}^{1.11} H^{0.24}} \ldots\ldots (13)$$

Pressure drop: the equation is used to design the pumping system of a cyclone or to determine the capacity of an existing cyclone.

$$P = \frac{1.88Q^{1.78} \exp[0.0055\phi]}{{D_c}^{0.37} {D_i}^{0.94} h^{0.28} ({D_u}^2 + {D_0}^2)^{0.87}} \ldots\ldots (14)$$

Sharpness of separation and classification: m in the equation (15) serves as a direct measure of sharpness of classification. Values of m>3 represent sharp classification while m<3 indicates relatively poor classification.

$$m = 1.94\exp[-1.58R_v\left(\frac{{D_c}^2 h}{Q}\right)^{0.15} \ldots\ldots (15)$$

Where: D_c = inside diameter of a hydrocyclone measured at the bottom of vortex finder

D_i = inside diameter of a hydrocyclone inlet or $(4A/\pi)^{0.5}$ for non-circular inlets

D_0 = inside diameter of the overflow or vortex finder of a hydrocyclone

D_u = inside diameter of the underflow, or apex, orifice of a hydrocyclone

h = free vortex height of a cyclone, which is defined as the distance from the bottom of the vortex to the top of the underflow orifice

H = pressure drop across a hydrocyclone expressed in head of slurry feed

Q = volumetric flow rate of hydrocyclone feed slurry

P = pressure drop across a hydrocyclone

ϕ = volumetric fraction of solids in the feed slurry

d = diameter of particle

ρ_s = solid density

ρ = density of liquid

R_v = recovery of feed volume to the underflow product

These equations show the independent effects and relative importance of all major variables which influence the operation of a hydrocyclone (Plitt, 1976).

CHAPTER 3

MATERIALS AND METHODS

The research was divided into four sections which included data collection, plant survey, flotation tests and the simulation of the primary milling circuit.

3.1 Plant Data Collection

Plant data was collected in order to understand the plant background, its current operations and the assays for the feed, concentrate and tailings. This information was helpful in providing a clear picture of the plant operations and in trouble shooting any problems that are linked to the research objectives. The following information was collected;

- Trojan concentrator flow sheet, mineralogical reports
- Plant operating parameters: feed rates, % solids in the ball mills, ball mill dimensions, mill speeds, particle size distribution of the primary mills feed, vortex spigot diameter, inlet diameter, vortex finder diameter, spigot diameter, cyclone diameter, cyclone feed density and pressure, flotation reagent dosages and target grind sizes.
- Daily Laboratory assay results of the concentrator's feed streams and concentrates.

3.2 Plant Survey and Analysis on Sampled Streams

The plant survey was carried out in order to identify particle classes that gave highest recoveries of valuable mineral and lower recoveries of MgO into the concentrate. Points at different stages of the flotation process where high MgO is recovered into the concentrates were also identified. The PSD and assays of the concentrates and tailings streams were carried out in order to understand the recovery mechanism of nickel and MgO into the concentrate. Samples from the primary and regrind cyclones overflow and underflow were also taken in order to assess their performance. The sample buckets were labelled prior to the survey and the sampling points and methods were communicated to the survey team. Samples from the selected streams as indicated in Table 3.1; were taken every hour, for four hours to obtain composite samples. The collected samples were dried, weighed and screened.

The heads sample and samples for each screen size were weighed, labelled and then sent for assaying. The plant operating conditions such as reagent additions, pulp density, air flow rates, impellor speed and the feed rate were also noted during the plant survey.

Table 3.1 Streams sampled during the plant survey

Sample Number	Process Stream Sample	Sample Number	Process Stream Sample
1	Rougher cells feed	17	Scavenger Final cleaner tailings
2	Rougher float bank concentrate	18	MgO removal cyclone feed
3	Rougher float bank tailings	19	MgO removal cyclone overflow
4	Rougher cleaner concentrate	20	MgO removal cyclone underflow
5	Rougher cleaner tailings	21	Regrind cyclone feed
6	Rougher Re-cleaner concentrate	22	Regrind cyclone overflow
7	Rougher Re-cleaner tailings	23	Regrind cyclone underflow
8	Rougher Final cleaner concentrate	24	Autokumpu Cell concentrate
9	Rougher Final cleaner tailings	25	Autokumpu Cell tailings
10	Scavenger Float bank concentrate	26	Final Tailings
11	Scavenger Float bank tailings	27	Final Concentrate
12	Scavenger Cleaner concentrate	28	Primary Cyclone feed
13	Scavenger Cleaner tailings	29	Ball Mill 1 discharge
14	Scavenger Re-cleaner concentrate	30	Ball Mill 2 discharge
15	Scavenger Re-cleaner tailings	31	Primary Cyclone underflow
16	Scavenger Final cleaner concentrate		

The sub-sieve (-32µm) particles of the rougher cells feed, final concentrate and final tailings streams were further analyzed using the cyclosizer at the University of Zimbabwe laboratories. Collected samples were then sent for assaying.

3.2.1 Cyclosizer Test Procedure

The sub-sieve particles (-32µm) were taken to the University of Zimbabwe for sizing. Twenty grams samples were pulped into a 250ml beaker using 50-100ml of water and the sample was placed in a cleaned sample holder.

The control valve was closed and the pump on the control panel was switched on. Air was bled from the cyclones by opening the apex valves of each cyclone. The control valve was fully opened and the sample was allowed to fully discharge for 5 minutes. The control valve was closed until it indicated the required elutriating flow rate and the elutriation time was then set at ten minutes. After ten minutes, the apex valves were opened and the samples were removed from the apex chamber into a beaker. The beakers were allowed to stand for twenty minutes and excess water was decanted. The samples were dried and weighed and the percentage of particles passing the 5^{th} cyclone was calculated. Tests were carried out at an elutriation flow rate of 12.11 litres per minute and at an average temperature of 22.5^{o}C. The relationship between the specified limiting particle separation size (d_i) and the effective particle separation size (d_e) is given by:

$$d_e = d_i . f_1 . f_2 . f_3 . f_4 \quad \ldots\ldots (16)$$

Where:

d_e = effective separation size of the cyclone

d_i = limiting particle separation size of the same cyclone at standard operating conditions

f_1, f_2, f_3 and f_4 = are correction factors for temperature, particle density, flow rate and elutriation time respectively (User Manual, 1981).

3.3 Simulation of the Primary Milling Circuit

Simulation was carried out, using the MODSIM simulator, on the primary milling circuit in order to come up with the optimum PSD for the coarse flotation circuit feed as indicated by plant survey results. The selection and breakage parameters of the ore were experimentally determined from the laboratory so that the milling circuit could be simulated. A Visual Basic Applications (VBA) excel based parameter estimation package was used the estimate the selection and breakage function parameters for the low grade ore. The hydrocyclone optimum operating parameters were also determined using the Plitt model equations.

3.3.1 Selection and Breakage Function Test

The sample with particle sizes ranging from 1700-106µm was prepared from the low grade ore and a sample of 1.030 kg was obtained.

A 300g sample was screened to obtain the initial PSD of the ore. An ore sample of 1.030 kg was placed in a 215 x 256mm ball mill, with 10kg of 30mm diameter steel balls. The fractional void filling (U), load volume fraction (J), mill speed fraction of critical speed (ϕ_c) and material feed characteristics (f_c) of the test mill are 0.25, 0.27, 0.75 and 0.027 respectively.

The following formulae were used to calculate: U, J and f_c

$$J = \frac{V_{load}}{V_{mill}} \ldots\ldots (17)$$

V_{load} = volume of load (volume of slurry + volume of ball)
V_{mill} = volume of the mill

$$U = \frac{V_{slurry}}{V_{voids}} \ldots\ldots (18)$$

V_{slurry} = volume of slurry at 70% pulp density
V_{voids} = volume of voids

It was assumed that volume of voids is 40% of the volume of the mill (Moys and Woollacott, 2014).

$$U = \frac{f_c}{0.4J} \ldots\ldots (19)$$ (researchgate.net, Accessed 08/08/15)

The mill was closed and fastened and placed on rollers and milling was carried out for 30 seconds (Moys and Woollacott, 2014). The mill was removed from the rollers and placed on the discharge table. Mill balls were cleaned using a brush and returned into the mill. The sample was removed from the mill, split using a spinning riffler and a 300g of the split material was screened to obtain the PSD of the mill discharge. The screening was carried out for 20 minutes using the sieve shaker and the screened material for the given size classes was weighed and recorded.

The screened as well as the rejected material were placed back into the mill. The procedure was repeated and the sample was milled a further 30, 60 and 120 seconds (metso.com, Accessed 8\10\2014).

3.4 Flotation Tests

Flotation tests were carried out using the CMC and guargum depressants in order to come up with the best, economic reagent suite that could reduce the MgO levels in the final concentrate. The depressants were benchmarked against the Betamin DZT 245 depressant that is currently used in the plant. The dosages of the CMC and guargum depressants, pH modifier used in the flotation tests were determined from other nickel plants reagent suites and average dosages were used (Bulatovic, 2007). The current plant dosages for the Betamin DZT 245 depressant, FTZ8 frother and SIPX collector were used. Different types of depressants behave differently and their strength also varies, hence their recommended dosages vary from one type of depressant to the other. Disseminated ore samples were collected from the mine with the help of a geologist for laboratory testing purposes. Plant return water was used in all the tests in order to simulate the pulp chemistry in the flotation plant.

3.4.1 Crushing and Milling

The ore was crushed using a laboratory jaw crusher and the sample passing -4mm was taken for milling. The spinning riffler was used to get a milling representative sample of 1.2kg.

The ore was placed in the rod mill, with a 15kg load of rods and 700ml of water. The sample was milled for 22 minutes to give a product of 52-55% passing 75µm. The laboratory rod mill was used instead of the ball mill because it produces a particle size distribution that is almost close to that obtained in closed-circuit ball mills (Wills and Napier-Munn, 2007). Batch grinding using ball mills produces a flotation feed with a wider PSD than that obtained in continuous closed-circuit grinding (Wills and Napier-Munn, 2007). All the samples were crushed and milled immediately before carrying out the flotation tests to avoid oxidation of liberated mineral surfaces.

3.4.2 Batch Flotation Tests

Ore from the laboratory rod mill was washed into a 3 litre cell and make up water was added to achieve the required pulp level. The pulp density was found to be 32% solids. Denver D12 flotation machines were used to carry out the flotation tests and one selected machine was used throughout all the tests. The impeller speed was set at 2000 rpm, based on Trojan laboratory procedure, and the pulp was conditioned for 8 minutes and 4 minutes after adding 50g\t $CuSO_4$ solution and 260g\t SIPX solution respectively. This was done to ensure that the pulp was thoroughly mixed with the reagents. A depressant and 3 drops of the FTZ8 frother were also added and the slurry was conditioned for 30 seconds after each addition. Pulp pH was also recorded and the $100cm^3$ heads samples was also taken before floating.

The impellor speed was then reduced to 1500 rpm and the air flow meter was opened slowly until the set point was reached (Eurus Mineral Consultants, 2012). Froth paddles were used to scrap the froth every 15 seconds while adding wash water to maintain the uniform pulp level (Eurus Mineral Consultants, 2012). Concentrates were collected into the concentrate trays at time intervals of 1, 5, 15, and 25 minutes (Eurus Mineral Consultants, 2012). The concentrates as well as the tailings sample were filtered and dried. Dried samples were weighed using an electronic balance and the masses were recorded, rolled using a sample roller, packed into sample sachets and sent for assaying. This method was used to screen the depressants by selecting the ones that gave the highest nickel recoveries, and low MgO content in the concentrate. The selectivity index, enrichment ratio and mass pulls obtained were also considered to evaluate the depressants.

3.4.3 Guargum Depressants

Three guargum depressants were first tested at the normal pH of around 8.9, and a single dose of 275g/t of the depressant was added. For the Betamin depressant, which is currently used in the plant, a 50g/t dose, equivalent to the current plant dose that was determined by the supplier, was added. Sodium carbonate was also added in order to raise the pH to values of around 10.2 at a dosage of 5000g/t. These depressants, though with different dosages, will be evaluated by comparing the reduction in MgO penalties and the cost of reagents per tonne of processed ore.

A test was also carried out without a depressant in order to understand the kinetics of both the valuable and floatable gangue minerals.

3.4.4 CMC Depressants

CMC depressants were tested using a dose of 360g\t and the pH of the slurry was raised from 8.9 to around 10.2 by adding 5000g\t of sodium carbonate.
Two tests were carried out at normal pH in order to assess the performance of CMC depressants at normal pH.

3.4.5 Collector Combinations

Different collector combinations were also tested with the standard Betamin depressant. The collectors that were blended with SIPX (sodium isopropyl xanthate) at a ratio of 1:1 were PNBX (potassium n-butyl xanthate), NC228 (2 mercaptobenzothiazole blend) and NC236 (mixture of potassium amyl xanthate (PAX) and sodium isobutyl xanthate (SIBX)). Concentrates were collected at the time intervals of 1, 5, 15 and 25 minutes in order to understand the kinetics of the valuable mineral and gangue minerals. The Betamin depressant was added before the test and after 1 minute into the flotation test in order to also investigate the effect of depressant stage addition on fast floating magnesia silicate gangue minerals.

3.4.6 Effect of Rejecting Iron Sulphide from the Concentrates

Two tests were carried out using a 1:1 blend of massive ore and disseminated ore. Batch flotation tests were carried out, one with copper sulphate and the other without copper sulphate in order to prevent the activation of iron sulphide. The concentrates were collected at 1, 5, 15 and 25 minute intervals.

CHAPTER 4

PLANT SURVEY AND MODSIM BASED SIMULATION RESULTS

The plant survey was carried out while the plant was running at an average solids feed rate of 90t\h, at a ball load of 30-35%, with the average nickel content of 1.21% and 33.41% MgO.

4.1 Plant Survey Results

The plant survey enabled us to get the overall indication of the plant operations and the results obtained were grouped according to their banks. Sub-sieve samples (-32µm) that were analysed by a cyclosizer are: rougher cells feed, final concentrate and final tailings. The calculated effective particle separation sizes for the cyclone system (d_e) were 27.5 µm, 19.2 µm, 12.5 µm and 9.2µm.

4.1.1 Rougher Cells Feed and Concentrate

The rougher cells feed had 54.7% of its particles less than 75µm and 34.7% of these particles were less than 32µm. Table 4.1; shows the head sample assays, the masses for each sieve size as well as their nickel and MgO assays.

Table 4.1: Rougher bank feed, concentrate and tailings assays

Screen size µm	Rougher cells Feed (g)	Ni %	MgO %	Rougher Bank Conc (g)	Ni %	MgO %	Rougher Bank Tails (g)	Ni %	MgO %
Initial Mass (g)	300			300			300		
Heads		1.21	33.41		2.88	30.32		0.49	31.34
425	1.5	0.38	32.06	0.54	0.16	17.99			
300	5.81	0.33	30.46	1.44	0.49	34.21			
212	18	0.38	31.16	5.96	0.47	34.03	0.8	0.26	32.4
150	29.55	0.64	31.52	18.94	0.9	33.51	7.65	0.4	33.64
106	39.86	0.94	32.65	28.7	2.09	31.36	38.32	0.43	33.99
75	41.13	1.39	34.15	40.62	3.18	30.81	60.34	0.46	33.89
53	33.02	1.58	35.56	40.21	3.96	28.69	70.99	0.49	31.11
45	13.57	1.58	35.59	24.99	3.8	27.12	41.92	0.53	29.5
32	13.59	1.84	35.37	22.45	4.19	27	6.18	0.63	29.65
-32	103.97	1.37	34.26	116.15	2.57	29.53	73.8	0.54	30.31

The target grind size was met with the feed assaying 1.21% Ni. Higher MgO contents above 35%, were recorded in the size ranges of 53µm to 32µm. These MgO particles are within the floatable particle size ranges and they need to be effectively depressed to reduce their recovery in the flotation circuit. The mass of particles that were below 32 µm, analysed using the cyclosizer was 98.55g. Table 4.2; shows the percentage mass of particles retained for each effective particle separation sizes; which indicates the smallest particle size that can be retained by the cyclone. Particle size of 27.5 µm had the highest nickel content of 4.92% and lowest MgO content of 18.16% and they constituted 5.71% of the sub-sieve particles. Higher MgO contents were found in the particle sizes of 19.2-9.2µm and they constituted 94.29% of the total sub-sieve particles. Particles that were below 10 µm had the lowest nickel content of 0.93% and they constituted 31.6% of the sub-sieve particles. This showed that 10.4% of the rougher cells feed had particles that are less than 10 µm. These can be recovered into the concentrate by entrainment or as slime coatings, while the other particle that are greater than 10µm are recovered by true flotation.

Table 4.2: Cyclosizer results expressed as the % of the initial mass of -32µm particles

d_e µm	Rougher Feed mass %	Ni %	MgO %	Final Conc mass %	Ni %	MgO	Final Tails mass %	Ni %	MgO %
27.5	5.71	4.92	18.16	19.4	18.81	2.72	9.18	0.61	27.63
19.2	29.67	1.55	29.12	27.81	14.37	9.14	30.59	0.31	31.41
12.5	33.01	1.1	29.98	27.1	9.64	15.29	31.28	0.28	31.56
9.2	31.61	0.93	30.33	25.67	7.32	18.46	28.95	0.3	31.16

The feed assaying 1.21% Ni and 33.42% MgO was upgraded to 2.88% Ni and 30.32% MgO. The rougher bank concentrate had 69.6% of its particles that were less than 75µm and 38.6% were less than 32µm. These fine particles had the highest recovery into the concentrate as shown in Table 4.1. Their nickel and MgO content were 2.57% and 29.53% respectively. Higher nickel grades were obtained on the intermediate particle sizes of 53µm to 32µm and they constituted 29.2% of the concentrate. The higher MgO content in particles that are above 75µm could have resulted from insufficient liberation of the particles.

The highest MgO content was recorded on the intermediate size particles of 75µm to 32µm, which are within the floatable size ranges. This shows that the rougher cells bank was not operating effectively or the depressant dosage added could have been insufficient.

The depressant additions are carried out manually in the plant and the dosages are adjusted accordingly after observing the colour of the froth and by visually estimating the grade of the feed. The operators can visually estimate the grade of the ore being fed into the mills by checking the colour of the ore. Massive ores have a dark grey colour with some golden shiny spots and the disseminated ores have the dull grey colour. The higher proportion of any of these ores determines the average grade in the plant and the presence of the golden colour in the froth generally indicates that the plant is processing high grade ores. On the other hand, when processing low grade ores which have a high talc content, the whitish froth instead of a grey froth in the first or second rougher cells can give an indication that we are floating more talc, hence depressant addition are increased accordingly. However, these visual estimates are not accurate since they are based on individual operator's discretion.

Generally, in different plant operations the depressant dosages differ with the MgO content in the feed, and the adjustment is made in line with the results given by the online stream analyser, which analyses the grade of the ore that gets into the milling circuit. The nickel and MgO assays would be of key interest since there is need to maximize and minimize their recoveries respectively. This helps reduce the reagent consumption in the plant and ensures that the optimum dosage of a particular depressant is not exceeded, especially with the guars, which can have a depressing effect on the valuable mineral as well. No online stream analyser is currently employed at Trojan concentrator plant. This leads to inconsistent depressant additions which vary with individual operators.

4.1.2 Scavenger Rougher Feed and Concentrate

The scavenger cells feed had 69% of its particles that were less than 75µm as shown in Table 4.3. The feed had less fine particles as compared to the rougher cells feed due to desliming that was carried out before regrinding the ore. Only 14.4% of the feed particles was less than 32µm and much of the feed was concentrated between the 106-53µm particle sizes. Particles that are less than 75µm had less MgO content as compared to those of the rougher cells feed. However, the higher nickel content in the feed was found in the particle sizes of 45µm to -32µm. Higher MgO contents were also obtained in particles that are above 75µm, due to insufficient liberation.

Table 4.3 Scavenger Roughers feed and Scavenger Roughers Concentrate

Screen size µm	Scav Roughers Feed (g)	Ni %	MgO %	Scav Roughers Conc (g)	Ni %	MgO %
Initial Mass	300			300		
Heads		0.79	32.81		1.87	27.59
300	0.57	0.34	34.54			
212	5.49	0.3	36.55			
150	26.22	0.33	35.22	1.89	3.34	27.07
106	60.48	0.45	34.99	22.22	3.35	28.19
75	89.04	0.76	33.56	70.42	2.81	27.15
53	47.94	0.97	31.69	63.75	3.23	22.8
45	21.99	1.29	30.27	27.59	3.52	19.67
32	5.11	1.64	30.09	28.59	4.61	19.44
-32	43.16	1.36	30.28	85.54	2.79	25.5

The scavenger rougher bank feed, that was deslimed, had less finer MgO particles and MgO content as compared to the rougher bank feed. This shows that desliming before the coarse floatation process could lead to reduced MgO slimes in the feed and in turn reduce the reagent consumption in the bank. The overflow obtained would be processed separately in the Outokumpu cell as well as the overflow obtained from the MgO removal/desliming cyclones as indicated in Figure 2.2, where nickel fines are recovered. Introducing the desliming unit could be beneficial since the desliming cyclones have low installation and operational costs. The overflow should be processed separately in the high intensity conditioning cell that is currently used in the plant. It will reduce the recovery of MgO due to slime coatings in the final concentrates and the recirculation of slimes in the flotation circuit.

The scavenger bank feed assaying 0.79% Ni and 32.81% MgO was upgraded to 1.87% Ni and 27.59% MgO. The scavenger bank concentrate had 68.5% of its particles that were less than 75µm as shown in Table 4.3. Fine particles that were less than 32µm had the highest recovery into the concentrate and 28.5% of these fine particles were recovered into the concentrate. They had higher MgO content as compared to other particle sizes that are below 75µm, but they were, however, lower than that of the rougher bank concentrates. The MgO content was reduced by 5.22% in the scavenger bank as compared to 3.09% in the rougher bank, indicating that desliming reduces the MgO content that is recovered into the concentrates.

4.1.3 Rougher and Scavenger Final Cleaner Concentrates

The rougher final cleaner concentrate assayed 14.07% Ni and 13.33% MgO and 78.9% of the particles were less than 75µm as shown in Table 4.4. The concentrate had 50.3% of its particles that were less than 32µm, which had the highest recovery into the concentrate. Higher MgO content was recorded in the particles less than 45µm. Particles that were greater than 45µm had lower MgO content.

The scavenger final cleaner concentrate had 85.9% of its particles that were less than 75µm and 36% of the particles were less than -32µm as shown in Table 4.4. The particles that were below 32µm had the highest recovery into the concentrate, however, they were less than those recovered in the rougher final cleaner concentrate.

They had the highest nickel content as compared to other particles that are less that 75µm. Particle sizes of -45µm to +32µm had lower MgO content as compared to those of the rougher final cleaners concentrate.

Table 4.4 Rougher final and scavenger final cleaner concentrates

Screen size µm	Rougher Final Cleaners Conc (g)	Ni %	MgO %	Scav Final Cleaners Conc (g)	Ni %	MgO %
Initial Mass (g)	300			300		
Heads		14.07	13.33		6.9	13.27
212	1.76	24.26	1.59			
150	8.85	23.34	2.24			
106	20.06	23.7	3.48	7.21	12.01	15.47
75	32.71	19.68	8.02	34.89	7.61	16.2
53	37.8	17.38	10.48	64.08	5.53	14.16
45	22.91	15.55	12.87	44.28	4.77	12.17
32	24.44	13.95	14.74	41.19	6.09	12.06
-32	151.47	13.58	14.53	108.35	8.28	12.65

Desliming the regrind mill feed led to reduced percentage of fines and MgO recovered into the scavenger final cleaner concentrate as compared to the rougher final cleaner concentrate. Higher MgO contents in particles greater than 53µm could have resulted from insufficient liberation of ore particles.

Regrinding is carried out before the scavenger flotation process in order to further liberate the valuable minerals from the gangue minerals. If the grinding process is not efficient enough, the recovery of the composite particles can reduce the grade of the concentrates.

4.1.4 Cleaner and Re-cleaner Banks

In the roughers section the feed was upgraded from 2.88% Ni and 30.32% MgO to give a re-cleaner concentrate assaying 7.72% Ni and 20.45% MgO. The scavenger cleaner bank upgraded its feed from 1.87% Ni to 27.59% MgO to give the concentrate of 3.45% Ni and 19.65% MgO. The scavenger cleaners feed had lower MgO content as compared to that of the rougher cleaners.

4.1.5 Outokumpu Cell Feed and Concentrate

The Outokumpu cell feed consisted mainly of very fine particles that are recovered as overflow of the desliming or MgO removal cyclone. Its feed had 49% of its particles which were less than 32µm. The feed and the tailings samples were difficult to screen due to high talc content in the samples. The fine feed that assayed 0.56% Ni and 33.93% MgO was upgraded to give a concentrate of 4.86% Ni and 26.13% MgO. The cell was operating efficiently and the MgO levels were reduced by 7.8%.

Table 4.5 Outokumpu cell concentrate

Screen size µm	Ok Cell Conc (g)	Ni %	MgO %
Initial Mass	300		
Heads		4.86	26.13
150			
106	4.4	1.43	31.42
75	34.6	3.2	29.57
53	41.17	3.41	28.07
45	24.86	4.03	26.46
32	13.17	4.76	26.07
-32	181.8	5.83	23.43

The Outokumpu cell concentrate had 60.6% of its particles which were less than 32µm and they had the highest nickel content of 5.83%, as shown in Table 4.5.

Intermediate particles of 75-45µm were also recovered into the concentrate. The Outokumpu concentrates are then fed into the scavenger cleaner bank to go through further cleaning. The recirculation of the Outokumpu concentrate into the scavenger cleaner circuit leads to the re-introduction of slimes into the flotation circuit. It would be beneficial to clean the Outokumpu cell concentrate in a separate cell and then discard the tailings. This will reduce the recovery of MgO into the final concentrate either by entrainment or due to slimes coating.

4.1.6 Final Concentrate and Final Tailings

The final concentrate consists of the mixture of the rougher final cleaner concentrate and the scavenger final cleaner concentrate. It had 81.1% of its particles which were less than 75 µm and 30% of the particles were less than 32µm as shown in Table 4.6. Higher nickel contents and lower MgO were found in particles that were greater than 75µm to 150µm, but they only contribute 18% of the final concentrate. Higher flotation rates are obtained in intermediate particle classes of 53-32µm, which also had higher MgO levels.

Table 4.6 Final concentrate and final tailings

Screen size µm	Final Conc (g)	Ni %	MgO %	Final Tails (g)	Ni %	MgO %
Initial Mass (g)	300			300		
Heads		11.08	13.98		0.27	32.2
212				5.31	0.27	32.95
150	2.2	16	6.15	22.46	0.28	33.64
106	10.9	16.87	9.02	41.86	0.25	34.41
75	43.57	13.37	12.09	58.34	0.22	33.31
53	71.54	10.15	13.98	57.18	0.24	31.74
45	40.79	10.24	14.44	25.44	0.28	32.19
32	38.25	10.72	12.98	12.82	0.31	31.41
-32	92.75	9.91	14.81	76.59	0.4	32.07

These need to be effectively depressed, to ensure that their recovery into the concentrate is reduced. The mass of particles which were less than 32 µm, analysed using the cyclosizer was 83.35g. Particle size of 27.5 µm had the highest nickel content of 18.14% and lowest MgO content of 2.72% as indicated in Table 4.2. These constituted 19.2% of the sub-sieve particles. Particle sizes of 19.2 and 12.5µm had the nickel grades of 14.37% and 9.64% respectively.

Higher MgO contents were in the particles sizes of 12.5 and 9.2 µm with the nickel assays of 15.29% and 18.46% respectively. These particles make up 52.77% of the sub-sieve particles in the concentrate. Particles which were below 10 µm had the lowest nickel content of 7.32% and they constituted 25.67% of the sub-sieve particles. Thus, 7.13% of the particles in the final concentrate were less than 10 µm.

It is assumed that particles that are greater than 10µm were recovered by true flotation into the concentrate, while the particles that are less than 10µm were recovered by entrainment (Runge, 2010). This indicates that most the magnesium silicates were recovered by true flotation, hence, the need to effectively depress them during the flotation process stages. The recovery by entrainment assessment was not carried out during this study, it can however be reduced by froth washing and using a moderate frother that allows froth drainage through bubble coalescence. The froth phase in the flotation process determines the grade of the concentrate that is produced.

The final tailings had 57.3% of its particles that were less than 75µm and 25% of these particles were less than 32µm as shown in Table 4.6. Sub-sieve particles had the highest nickel content of 0.4%. Higher nickel losses to the tailings were realised in the finer size particles of 32 to -32µm. Hence, reducing overgrinding in the mills will reduce the percentage nickel loss into the tailings. The mass of particles that were less than 32 µm was 70.91g. Particle size of 27.5 µm had the highest nickel content of 0.61% and lowest MgO content of 9.18% as shown in Table 4.2. They constituted 9.18% of the sub-sieve particles. Higher MgO contents were in the particles sizes of 19.2 and 12.5µm with the nickel assays of 0.31% and 0.28% respectively. Particles that were below 10 µm had the nickel content of 0.3% and they constituted 28.95% of the sub-sieve particles. This showed that 6.84% of the final tailings had particles that were less than 10 µm.

4.2 Selection and Breakage Function Test Results

A VBA excel based parameter estimation package was used to determine the selection and breakage functions from the laboratory test results that were obtained.

4.2.1 Selection and Breakage Function Parameters

The selection function model parameters as shown in Equation (8); kappa (∧) and alpha (α) were 0.000233052 and 0.98379289 respectively. The breakage function parameters as shown in Equation (10); phi (ϕ), gamma (γ) and beta (β) were 0.22177735, 29.95900904 and 0.618380027 respectively. These parameters enable us to calculate the selection function values for all size intervals in the feed and the expected size distribution in each particle class of the ore that results from the breakage. After scale up, the selection functions at 1mm for ball mills 1 and 2 were found to be 0.27279 and 0.28286 respectively. The scale up Equation (11); was used to calculate the milling rates of the plant mills from the laboratory milling test results. These results were used in the simulation of the milling circuit using the mill model.

4.2.2 The Selection Function of the Top Size Particle Class of Low Grade Ores

The rate of milling for the top size particle class (-2000µm to 1700µm) of the disseminated ores was found to be 0.243 as indicated by the gradient in Figure 4.1. These selection function values are a function of ball size, fraction of voids filled and load volume fraction of the mill.

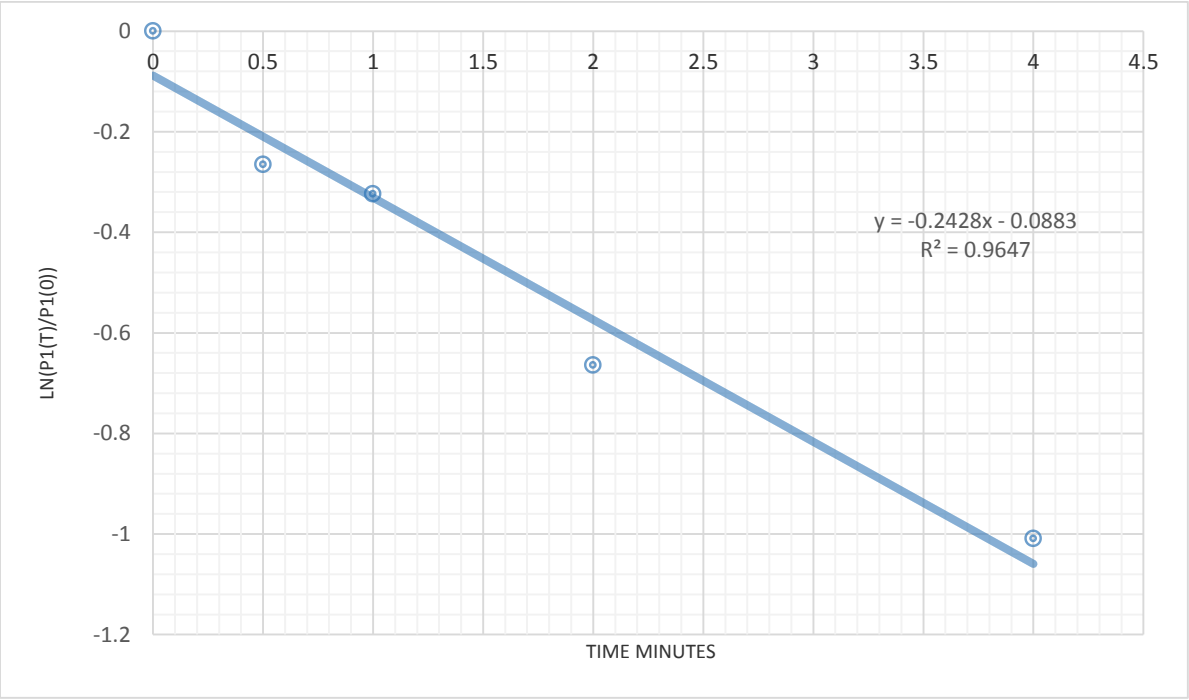

Figure 4.1 Selection function of top size particle class of the disseminated ores

The S_1 value of the same particle size estimated by the parameter estimation package was found to be 0.29345. A reasonable linear fit was applied to the experimentally derived S_1 value with the R^2 value of 0.9647. The disseminated ore sample assayed 0.63% Ni, 40.0% MgO and 6.51% Fe. The low milling rate of the 1700µm particles indicates that the low grade ore is fairly hard. However, some fines could be seen in the low grade samples in the form of talc. The selection functions obtained using the parameter estimation package, increased with the increase in particle size. The selection function increases with increase in particle size up to a certain particle size, which gives the maximum grinding rate, then it begins to drop (Moys and Woollacott, 2014).

4.2.3 Selection Function of the Top Size Particle Class of Massive Ores

The rate of milling for the top size particle class (-2000µm to1700µm) for the massive ores was found to be 0.9995 as indicated by the gradient in Figure 4.2.

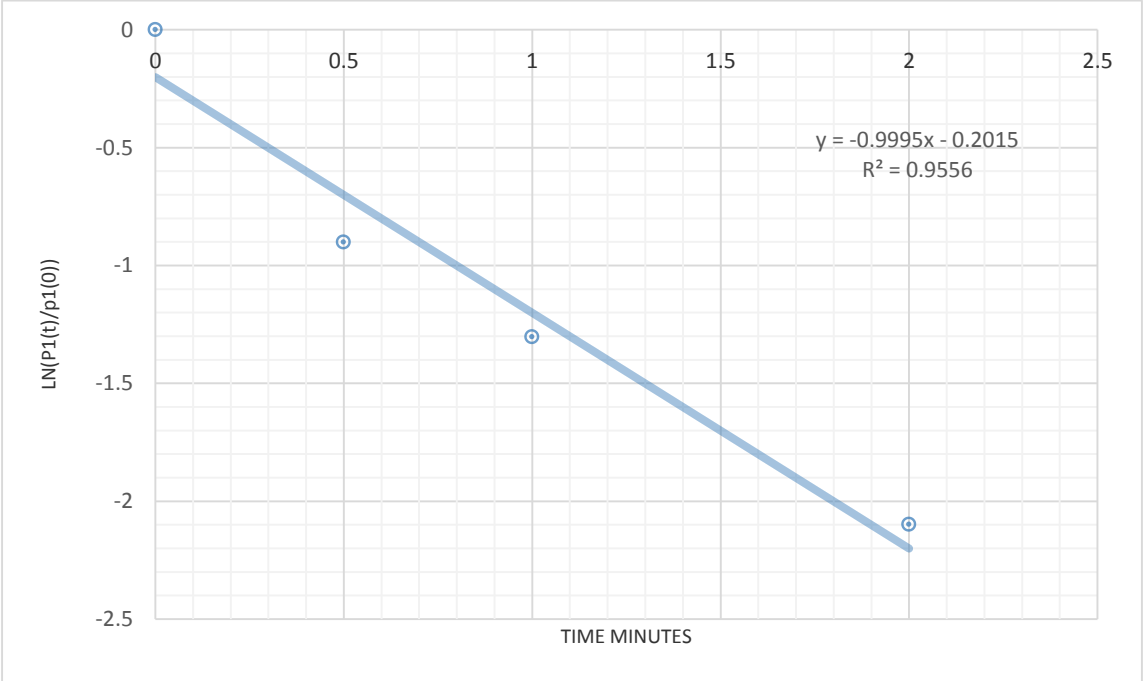

Figure 4.2 Selection function of top size particle class of the massive ores

The massive ore sample assayed 12.49% Ni, 0.57% MgO and 46.58% Fe. The rate of milling for the top size particles of the massive ores was found to be at least three times higher than that of disseminated ores under the same milling conditions.

Hence, more fines are most likely to be generated when the mill feed is blended with high proportions of massive ores. Experimental work has shown that the first-order model holds even when grinding occurs in the presence of other particle classes (Moys and Woollacott, 2014).
It was concluded that the presence of other particle classes did not affect the rate of grinding of any particular class of particles (Moys and Woollacott, 2014).

From this finding, it can be deduced that increasing the ore feed rate when milling high grade ores will reduce the production of excessive fines. However, optimum milling conditions still need to be determined using a simulator for a range of high grade feed assays, from 1.7% -4% Ni. The milling rate will differ due to the different blending ratios for different grades of the mill feed. There is no fixed blending ratio that is currently applied at the moment and this results in a large variation in the grade of the mill feed. Reducing the ball load when milling the high grade ore will also help reduce the fines produced. Using a professional simulator that allows the variation of more milling parameters will enable the attainment of more accurate results as compared to the MODSIM academic version.

As the grade of the mill feed increases the ore feed rate in the plant is reduced in order to achieve high recoveries in the flotation circuit. However, this creates the problem of generating more fines due to increased residence time in the ball mills. The solution would be to increase the feed rate when running with the high grade ores and reject the iron from the concentrate by taking out the activator from the reagents suite. High percentages of iron, as high as 38% have been recovered into the concentrates, at feed grades with nickel content greater than 3%. High iron recoveries in the concentrate also create problems of chocking pumps at the thickener.

4.3 Primary Milling Circuit Simulation Results

The MODSIM simulation results show that at a feed rate of 100t\h, low grade ore ranging from 0.6 to 1.0% nickel, could be milled at a ball load of 35% to obtain 55-60% passing 75µm as the primary cyclone overflow. Figure 4.3 shows the simulation results with their stream fly outs.
The target grind size could be achieved when milling low grade ores without producing excessive fines in the milling circuit. The cyclones must be operated at near roping conditions in order to minimize the amount of fines recovered to the underflow.

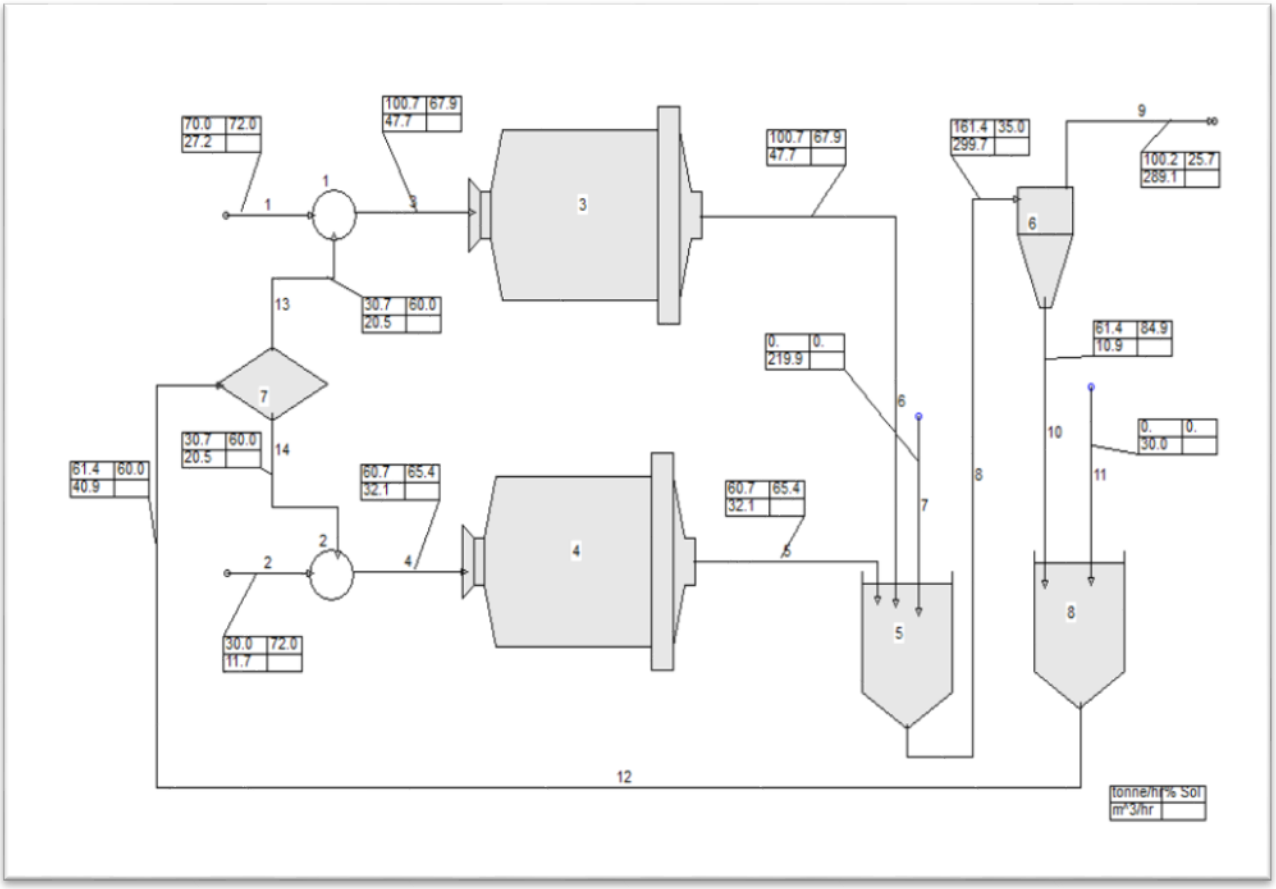

Figure 4.3 Simulation results of the Trojan's primary milling circuit

The PSD of the cyclones feed, underflow and overflow are shown in Figure 4.4. Overgrinding with this type of ore is unlikely to take place except in situations where the spigots are excessively worn out, which increases the percentage of fines recovered into the cyclone underflow. Factors which affect the performance of the hydrocyclone are: operating pressure, vortex and spigot diameters, feed size PSD, pulp density and slurry viscosity.

The cyclone pressure or feed rate and feed density can be adjusted to obtain the required cut point for different feed size distributions in the circuit. Particles recovered to the underflow which were less than 86µm increased from 3.95% to 5.66% when the spigot diameter was increased from 110mm to 120mm. The mass of these particles increased from 2.45 to 4.49 tonnes per hour. From the plant survey results it has been noted that 11.14% of the particles that are less than 75µm were recovered into the primary cyclone underflow. The underflow had 7.2% of the particles which were less than -32µm and these particles had relatively lower nickel content and high MgO content.

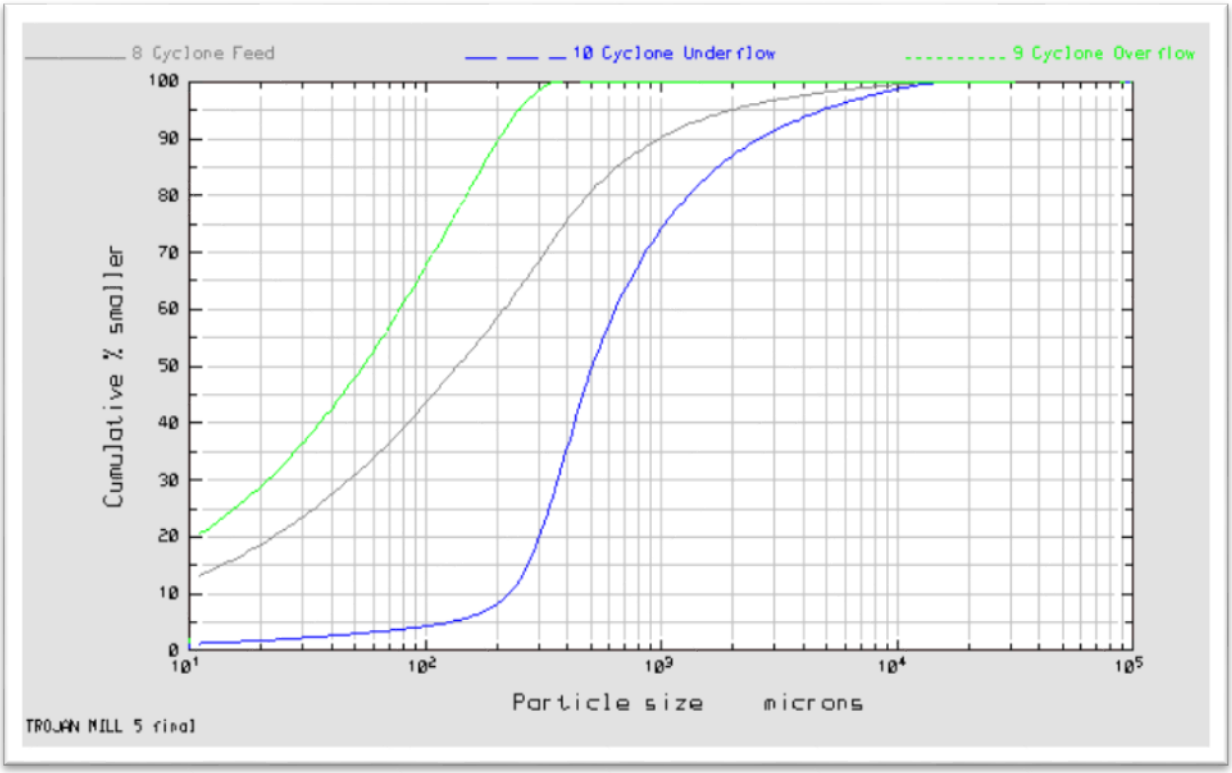

Figure 4.4 Particle size distribution of the primary cyclones feed, underflow and overflow

The high recovery of fine particles into the underflow could have resulted from excessive wear of spigots. The maximum wear allowed in the spigots is 10mm, and they need to be monitored and replaced when they reach the maximum diameter to reduce the recovery of fines into the underflow. Maximum spigot wear allowed should also be reduced when milling high grade ores.

4.3.1 Effect of Varying the Pulp Density of the Primary Cyclone Feed

The primary cyclone feed pulp density ranges from 26%-36% solids, depending on the feed rate and grade of the ore. The pulp density of the ore was varied from 25%-35% solids. It was observed that the cut size increased with increase in pulp density of the slurry. The pulp density of the underflow also decreased with a decrease in the cyclone feed pulp density. However, the mass of particles recovered into the underflow increased from 3.42% passing 86µm at pulp density of 30% solids to 3.95% passing 86µm at pulp density of 35% solids. Operating cyclones at high pulp density results in coarser overflow caused by the increase in drag force on each particle (wordpress.com, Accessed 10/08/15).

At a pulp density of 25% solids, the recovery to the underflow was increased to 3.7% passing 86µm as compared to 3.42% passing 86µm at pulp density of 30% solids. Lower pulp densities allows smaller and lighter particles to sink, resulting in reduced cut sizes. The optimum pulp density of the primary cyclone feed, which gave the lowest recovery of fines into the underflow, was 30% solids and gave a cut size of 192.9µm. Hydrocyclones should be operated at lowest possible density, while achieving the required separation size and overflow density (wordpress.com, Accessed 10/08/15).

CHAPTER 5

FLOTATION TESTS RESULTS

Flotation tests were carried out to screen the depressants and collectors. Stage addition of the depressant was also investigated by adding another depressant dose after 1 minute into the flotation test.

5.1 Depressant Screening Test Results

The Eurus Mineral Consultants laboratory float procedure was used as a guide to carry out the batch flotation tests. The impellor speed (rpm) and air flowrate that corresponded to the 3 litre cell used in the flotation tests were read from the Denver D12 laboratory flotation machine chart (Eurus Mineral Consultants, 2012). The depressants were evaluated based on their nickel recovery, enrichment ratio, mass pull and their selectivity index. The Goudin selectivity index between two minerals was applied (Kiyotsuno, 1963).

$$SI = \sqrt{\frac{N}{n} \times \frac{m}{M}} \quad \ldots\ldots (20)$$

Where

N, n are % nickel in concentrate and tailings respectively

M, m are % magnesium oxide in concentrate and tailings respectively

If selectivity index is 1, it is an indication that there is no separation between two minerals and if the separation is perfect the selectivity index is infinite (Kiyotsuno, 1963).

The target nickel recoveries are set by the plant for a given nickel content in the feed. They range from 65.69%-80.01% for nickel feed grades of 0.54%-1.2%.

The recommended mass pulls in the base metal flotation circuits, for each stage process, ranges from around 10-15% to give optimum results (allmetallurgist.com, Accessed 10/10/15).

The nickel grade is the measure of the nickel concentration in the concentrate and the enrichment ratio measures the extent to which the feed has been upgraded. The higher the grade or the enrichment ratio, the greater the upgrade.

5.1.1 Guargum Depressants

Table 5.1 shows the flotation test results obtained using the guargum depressants. The nickel and MgO recoveries, selectivity index, enrichment ratio and mass pull for the test results were compared. The depressants were tested at normal pH around 8.95 and at a raised pH of 10.2.

5.1.1.1 Tests at Normal pH (8.95)

At pH 8.95, the DLM RS and DLM PDE depressants had better selectivity as compared to the standard Betamin depressant. They had higher nickel recoveries, enrichment ratios, nickel grades and lower MgO recoveries into the concentrate as compared to the standard Betamin depressant.

DLM RS had the selectivity index of 6.23 and its nickel recovery was 10.96% above the set plant target recovery. Its mass pull was 8.98% and its enrichment ratio was 8.29. It had the nickel concentrate grade of 4.72% and the MgO recovery of 7.18% into the concentrate. The lower mass pull of the DLM RS depressant resulted in the higher nickel grade in the concentrate and at the same time gave good nickel recovery. This shows that the depressant is more selective towards the gangue minerals as compared to the other depressants.

DLM PDE had the selectivity index of 5.78 and its nickel recovery was 5.40% above the set plant target recovery. Its mass pull was 11.20% and its enrichment ratio was 6.72. It had the nickel concentrate grade of 4.10% and the MgO recovery of 8.20% into the concentrate. DLM PDE gave results second to the DLM RS and it had a moderate mass pull and gave a good nickel recovery. The increased mass pull resulted in lower nickel grades as compared to that of the DLM RS depressant,

The standard Betamin depressant had the selectivity index of 5.57 and its nickel recovery was 5.07% above the set plant target recovery.

Its mass pull was 15.30% and it was higher than that of other depressants, which could have resulted in reduced selectivity against MgO. This also resulted in lower nickel grades as shown in Table 5.1. It had the nickel concentrate grade of 3.21%, enrichment ratio of 5.26 and the MgO recovery of 10.97% into the concentrate.

Table 5.1 Guargum depressants flotation test results

Depressant	Betamin (std) pH 8.95	Betamin pH 10.2	No Depressant pH 8.95	Cytec S9349 pH 8.95	Cytec S9349 pH 10.2	DLM PDE pH 8.95	DLM PDE pH 10.2	DLM RS pH 8.95	DLM RS pH 10.2
Ni Recovery %	74.55	75.49	86.90	66.97	69.87	74.88	68.39	78.36	69.16
Mass of feed (g)	1147.72	1151.15	1142.98	1155.23	1155.76	1147.07	1154.05	1154.12	1154.05
Mass of concentrate (g)	162.55	173.60	201.09	136.19	152.39	127.85	80.55	109.11	86.71
Ni grade in feed %	0.61	0.63	0.54	0.68	0.68	0.61	0.64	0.57	0.67
Ni grade in Concentrate %	3.21	3.15	2.67	3.86	3.60	4.10	6.27	4.72	6.17
Ni grade in tails %	0.14	0.15	0.14	0.24	0.26	0.17	0.17	0.16	0.23
MgO grade in feed %	41.72	41.84	41.41	42.60	41.87	41.00	39.83	39.65	41.29
MgO Recovery %	10.97	11.70	14.20	8.25	9.74	8.20	4.69	7.18	5.48
MgO grade in concentrate %	32.32	32.45	33.43	29.83	30.93	30.17	26.74	30.12	30.09
MgO grade in tails %	43.67	43.64	42.95	44.98	45.85	41.88	41.27	39.63	42.15
Selectivity Index	5.57	5.32	4.95	4.93	4.53	5.78	7.55	6.23	6.13
Enrichment Ratio	5.26	5.01	4.94	5.68	5.30	6.72	9.80	8.29	9.20
Mass pull %	15.30	15.98	15.83	12.14	12.56	11.20	7.70	8.98	7.41
Ni Target Recovery %	69.48	70.43	65.69	72.61	72.61	69.48	70.89	67.40	72.20
Variance	5.07	5.06	21.21	-5.64	-2.74	5.40	-2.50	10.96	-3.04

Cytec S9349 had poor results as compared to the other guargum depressants. Its nickel recovery was 5.64% below the target plant recovery and it had the lowest selectivity index of 4.93. This shows that the reagent had a depressing effect on the valuable mineral, leading to lower nickel recovery. Its nickel grade and enrichment ratio was however, higher than that of the Betamin depressant. The total MgO recovered into the concentrate was also lower. Thus, the Cytec S9349 was disqualified on the basis of its lower selectivity index that resulted from its lower nickel recovery into the concentrate.

5.1.1.2 Tests at pH 10.2

All the depressants had reduced selectivity when they were tested at elevated pH except for the DLM PDE depressant as shown in Table 5.1. The selectivity indices of DLM PDE and DLM RS were fairly good but their nickel recoveries were also reduced. They had the selectivity index of 7.55 and 6.13 respectively. The nickel recoveries were 2.50% and 3.04% below the set target. They had low mass pulls of 7.70% and 7.41% respectively. Nickel grades obtained were higher and adding $NaCO_3$ increased selectivity of the depressant, but that resulted in reduced nickel recovery into the concentrate. Their mass pull values were below the recommended range of 10-15% that gives optimum results in the flotation of base metals (allmetallurgist.com, Accessed 10/10/15).

5.1.2. CMC Depressants

All the CMC depressants, tested at pH 10.2, gave nickel recoveries which were below the target plant levels except for the ND 522 depressant as shown in Table 5.2. ND 522 had a fairly good selectivity index of 5.59 and the nickel recovery that was 2.25% above the set target recovery. Its mass pull was 13.49% and its enrichment ratio was 5.79. It gave the results that were slightly better than those of the standard Betamin depressant. The use of the ND 522 depressant could not be justified, since it comes with the added cost of using the pH modifier

The selectivity indices of the other CMC depressants were lower than that of the standard Betamin depressant. This could have been caused by their lower nickel recoveries and the slightly higher nickel content in the tailings. ND521 had the highest selectivity index of 5.84 but it had the lowest nickel recovery; 13.97% below the target recovery.

Table 5.2 CMC depressants flotation test results

Depressant	Depramin 267 pH 10.2	Depramin 347 pH 10.2	Depramin 177 pH 10.2	ND 523 pH 10.2	ND 522 pH 10.2	ND 521 pH 10.2	Depramin 267 pH 8.95	ND 521 pH 8.95
Ni Recovery %	70.74	66.24	69.69	71.06	74.45	61.55	77.50	62.21
Mass of feed (g)	1153.89	1160.86	1158.81	1147.52	1158.64	1160.72	1130.82	1153.68
Mass of concentrate (g)	132.24	110.14	131.38	154.07	149.11	141.84	150.06	175.30
Ni grade in feed %	0.70	0.74	0.69	0.67	0.67	0.76	0.61	0.71
Ni grade in Concentrate %	4.32	5.17	4.24	3.55	3.88	3.83	3.56	2.91
Ni grade in tails %	0.25	0.26	0.24	0.17	0.17	0.16	0.23	0.18
MgO grade in feed %	42.88	41.53	42.35	40.39	40.66	39.44	41.98	41.65
MgO Recovery %	8.31	6.71	8.04	10.22	9.73	9.03	10.20	11.79
MgO grade in concentrate %	31.10	29.36	30.02	30.76	30.73	29.15	32.26	32.33
MgO grade in tails %	43.69	43.54	42.97	41.47	42.13	41.62	42.72	42.96
Selectivity Index	4.93	5.43	5.03	5.30	5.59	5.84	4.53	4.63
Enrichment Ratio	6.17	6.98	6.15	5.29	5.79	5.04	5.84	4.09
Mass pull %	11.05	9.78	11.25	14.81	13.49	16.36	11.40	19.43
Ni Target Recovery %	73.40	74.85	73.01	72.20	72.20	75.52	69.48	73.78
Variance	-2.66	-8.61	-3.32	-1.14	2.25	-13.97	8.02	-11.57

Depramin 267 and ND521 were also tested at normal pH of around 8.95. The results in Table 5.2; show that their selectivity was reduced by 0.4 and 1.21 respectively. The enrichment ratio was also reduced while the MgO content recovered into the concentrates was increased. This indicates that the CMC depressants were less effective at normal pH.

5.1.3 A Comparison of MgO-Nickel Recovery for Betamin, DLM RS and DLM PDE Depressants

Figure 5.1; shows the MgO-nickel recovery curves for DLM RS, DLM PDE and the Betamin standard depressant. Betamin depressant had the highest MgO recovery into the concentrate, it had a good depressing effect at the beginning of the test, but it was however short-lived. This led to higher MgO recoveries in the successive concentrates that were collected.

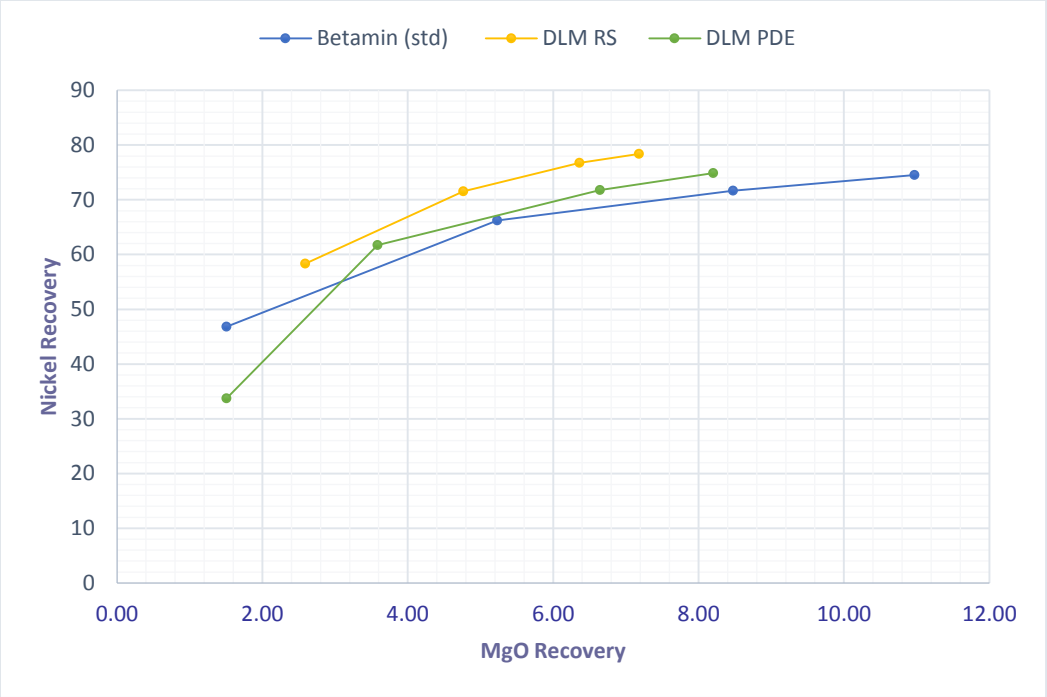

Figure 5.1 MgO-Nickel recovery curves of Betamin, DLM RS and DLM PDE depressants

The depressing effect of DLM PDE and DLM RS was fairly good throughout the test, although the MgO recovery for the DLM RS depressant was higher during the first minute of the test. It then maintained a good depressing effect throughout the test and this resulted in the lower MgO recovery into the concentrate as shown in Figure 5.2.

The DLM PDE depressant had a good depressing effect at the beginning of the test, and its effect slowly faded after 15 minutes into the flotation test. Hence, its MgO recovery into the concentrate was higher than that of the DLM RS depressant. Flotation test results showed that DLM RS and DLM PDE had better selectivity towards MgO as compared to the Betamin depressant that is currently used in the plant.

5.1.3.1 Effect of Varying Dosages of Betamin, DLM RS and DLM PDE Depressants

Further tests were carried out, at 130g/t, using Betamin, DLM RS and DLM PDE depressants and the results obtained are given in Table 5.3. This investigated the effect of increasing the dosage of the Betamin depressant, from 50g/t-130g/t, as well as reducing the dosage of the DLM RS and DLM PDE depressants, from 275g/t-130g/t.

Table 5.3 Effect of varying dosages of Betamin, DLM RS and DLM PDE depressants

Depressant	Betamin (std) 50g/t	DLM PDE 275g/t	DLM RS 275g/t	Betamin 130g/t	DLM PDE 130g/t	DLM RS 130g/t
Ni Recovery %	74.55	74.88	78.36	71.24	70.29	84.65
Mass of feed	1147.72	1147.07	1154.12	1134.24	1146.39	1139.25
Mass of concentrate	162.55	127.85	109.11	137.09	163.32	184.96
Ni grade in feed %	0.61	0.61	0.57	0.72	0.54	0.75
Ni grade in Concentrate %	3.21	4.1	4.72	4.24	2.66	3.91
Ni grade in tails %	0.14	0.17	0.16	0.24	0.24	0.22
MgO grade in feed %	41.72	41	39.65	40.91	41.19	40.66
MgO Recovery %	10.97	8.2	7.18	8.93	10.7	12.33
MgO grade in concentrate %	32.32	30.17	30.12	30.23	30.94	30.88
MgO grade in tails %	43.67	41.88	39.63	42.27	42.42	42.93
Selectivity Index	5.57	5.78	6.23	4.97	3.9	4.97
Enrichment Ratio	5.26	6.72	8.29	5.89	4.93	5.21
Mass pull %	15.3	11.2	8.98	11.99	12.37	14.36
Ni Target Recovery %	69.48	69.48	67.4	74.15	75.19	74.5
Variance	5.07	5.4	10.96	-2.91	-4.9	10.15

Increasing the Betamin depressant reduced its selectivity index from 5.57 to 4.97 as shown in Table 5.3. The nickel recovery was reduced from 5.07% above the target recovery to 2.91% below the target recovery. The MgO recovery into the concentrate was reduced from 10.97% to 8.93% and the mass pull was also reduced from 15.3 to 11.99. Nickel grade was increased from 3.21 to 4.24%.

Increasing the Betamin depressant dosages, reduced the MgO content recovered into the concentrate, however, it also had a depressing effect on nickel as well, as indicated by the reduced nickel recoveries.

DLM PDE had lower MgO recovery and lower selectivity index as compared DLM RS at a lower dosage, however, its recovery was 4.9% below the target recovery. The selectivity index was reduced from 5.78 to 3.9. The MgO recovery was increased from 8.2% to 10.7% and the enrichment ratio was reduced from 6.72 to 4.93. The concentrate grade was reduced from 4.1 to 2.66%. It had a depressive effect on nickel at a lower dose of 130g/t.

Reducing the DLM RS depressant dose, reduced the selectivity index from 6.25 to 4.97. The enrichment ratio was reduced from 8.29 to 5.21, and the mass pull was increased from 8.98 to 14.36. MgO recovery was increased from 7.18% to 12.33%. There was no change in nickel recoveries obtained against the set target, indicating that the depressant has no depressive effect on nickel at higher and lower dosages. It has an advantage over the other depressants of selectively depressing the magnesium silicates.

DLM RS depressant gave the best results at a dosage of 275g/t, with highest nickel and lowest MgO recovery into the concentrate. The depressant gave a lower mass pull and higher nickel recovery and grades. The mass of the concentrate collected was 109.11g, while that collected using the Betamin depressant was 162,55g. A 32% reduction in the mass of concentrates will also reduce transport costs. However, the given results show that the depressant is only effective at higher doses as compared to the Betamin depressant. An optimum dosage that lies between 130g/t and 275g/t still need to be determined. These flotation results have been benchmarked with the depressant that is currently used in the plant and this gives an indication that they can perform well in the plant.

The batch flotation test results are a good measure of pulp floatability and a poor indicator of the full-scale froth phase performance (Runge, 2010). When using the mechanical cells, it is assumed that the pulp selectivity in the small scale cell is similar to what would be achieved in the full-scale flotation machine (Runge, 2010). Hence, a plant trial should be carried out to verify the results that were obtained while using the DLM RS depressant. The optimum dosage for this depressant still needs to be determined before the plant trial.

The overall MgO reduction in the concentrates will be determined during the plant trial, to assess if the objective of reducing it to levels below 12% has been met. MgO levels were reduced by 3.79% during the roughing stage while using the DLM RS depressant and a further reduction is expected during the cleaning stages. However, using this depressants, will come at a cost of increased dosages per tonne of ore as compared to the Betamin depressant. The cost benefit of reduced MgO penalties, increased nickel recovery and reduced transport costs needs to be weighed against the cost of increased depressant consumption per tonne of ore after the plant trial.

On the other hand, the MgO which is currently treated as an unwanted penalty element can be beneficiated and sold as a by-product of the Trojan mine concentrator. It is used in the refractory industry to make the magnesite bricks, to produce cement, crucibles etc. The MgO in tailings can be beneficiated, through further flotation of the tailings to obtain an MgO concentrate. This excludes the cost of milling, which is usually a major cost in the mineral beneficiation process.

5.1.4 Test with no Depressant

Figure 5.2; shows the difference in nickel grades and recoveries after carrying out a test using the standard Betamin depressant and a test without a depressant. It shows the nickel recovery against the concentrate grade at different time intervals. Carrying out the flotation test without a depressant increased the recovery of both the valuable mineral and magnesium silicates, but decreased the nickel grade in the concentrate. A much larger volume of the first concentrate was collected when the test was carried out without a depressant, indicating that the collector used activated the magnesium silicates as well.

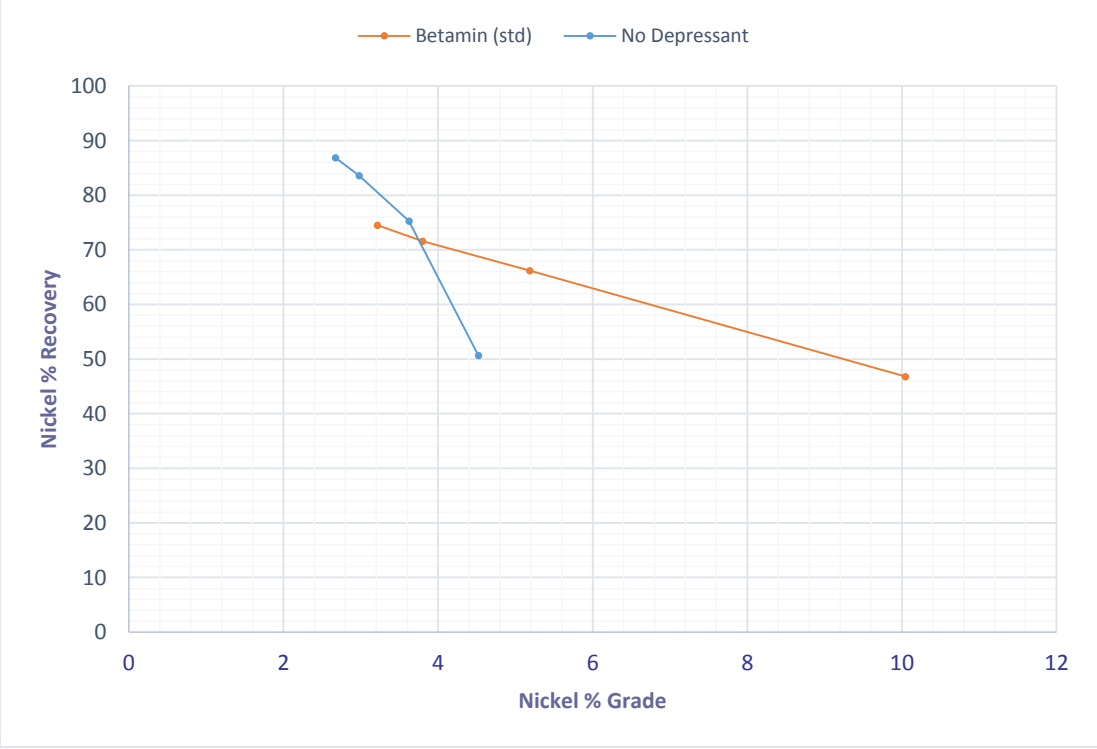

Figure 5.2 Nickel grade-recovery curves for tests carried out without a depressant and with a Betamin depressant

The mass of the concentrate collected after the first minute without using a depressant was 69.22g while that collected after adding the depressant was 32.6g. This indicates that the mass pull was far much higher during the first minute when no depressant was added. The first concentrate had low nickel grade of 4.51% and very high MgO content of 28.21% as compared to that of 10.04% Ni and 22.11% MgO when a Betamin depressant was added. The selectivity index was also increased from 4.95 to 5.57 after adding the Betamin depressant as shown in Table 5.1. The mass pull was reduced from 15.83% to 15.30% after adding the depressant.

5.1.5 Flotation rates of MgO after adding a Betamin depressant and with no depressant

A depressant is added to lower the flotation rate constant of the unwanted gangue minerals in the flotation cell. The flotation rate constant of MgO (kMgO) was found to be 0.0044min^{-1} when no depressant was added as shown in Figure 5.3. The total MgO recovery into the concentrate was 14.2%. The MgO recovery at 1, 5, 15 and 25 minutes was 4.13%, 4.07%, 3.70% and 2.31%. The highest MgO recovery was realised during the first minute of the flotation test followed by one recovered after 5 minutes.

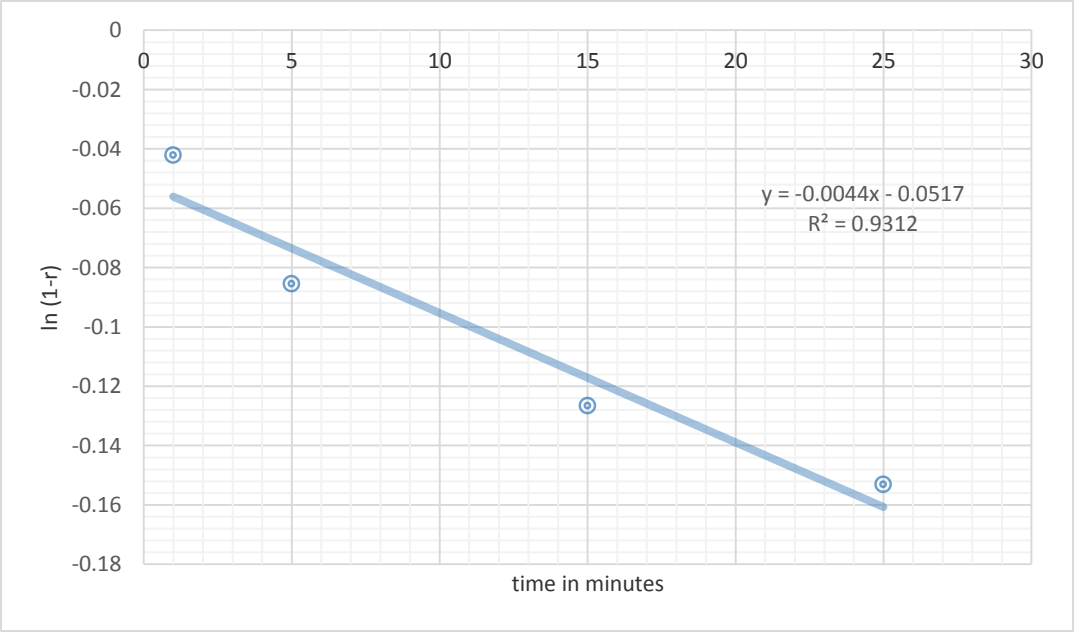

Figure 5.3 The MgO rate constant for a test carried out without adding a depressant

This indicates that most of the MgO is fast floating, since 57.7% MgO was recovered within the first five minutes of the test. Slow floating magnesium silicates, 6.01% MgO, were recovered between 6-25 minutes of the test, while 8.2% MgO was recovered during the first 5 minutes of the test.

The flotation rate constant of MgO (kMgO) was found to be $0.0039min^{-1}$ when a Betamin depressant was added as shown in Figure 5.4. Adding a single dose of the depressant reduced kMgO by $0.0005min^{-1}$. The overall MgO recovery into the concentrate was 10.97%, giving an overall reduction of 3.23%.

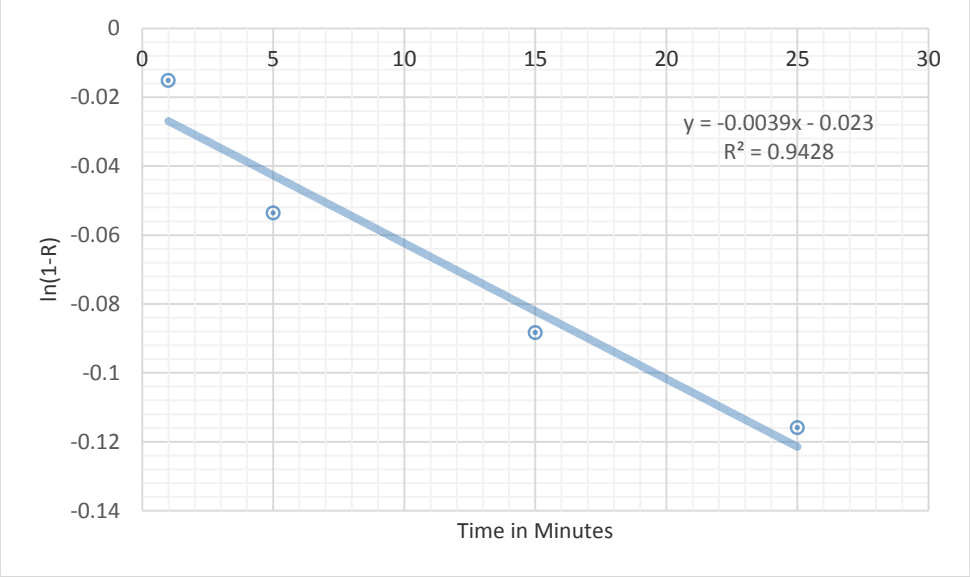

Figure 5.4 The MgO rate constant after adding a Betamin depressant

The flotation test carried out without a depressant indicated that most of the MgO was recovered during the first five minutes of the test. With an understanding of the kinetics of the magnesium silicates obtained, it could be deduced that the most of the MgO was fast floating, hence, the need to depress it effectively during the early stages of the process. The fast floating magnesium silicates when recovered into the rougher bank concentrates will still have greater chances of being recovered in the cleaner banks. Hence, reducing their recovery into the rougher bank concentrate will in turn reduce their recovery into the final concentrates.

The general difference between the MgO content in the concentrates collected after 1 minute and after 5 minutes was noted in all the tests carried out. It varied from around 9% to 14% indicating that another depressant dose may need to be added during the early stages of the flotation process, to further supress the fast floating magnesium silicates in the ore. It has been assumed that the fast floating MgO particles were recovered by true floatation. However, the magnesium silicates can be recovered into the concentrate by true flotation, entrainment, as slimes coatings and as composite particles (Pietrobon et al., 1997). Using an ASPEX scanning electron microscope to do a mineral identification and liberation scans for the samples that are collected at different time intervals, will enable the recovery mode of MgO to be fully quantified. Necessary adjustments in the milling circuit based on the liberation data obtained can then be implemented.

5.1.6 Effect of Adding another Depressant Dose after One Minute

After observing that most of the MgO was fast floating, a test was carried out, whereby another 50g/t Betamin depressant dose was added after 1 minute into the flotation test. It was noted that the second concentrate that was collected after 5 minutes had much higher MgO contents for most depressants that were tested, indicating that their depressing effect was short-lived, with the exception of DLM RS and DLM PDE depressants. The second dose was added in order to depress the fast floating MgO as much as possible. The Betamin depressant is currently added in two stages in the rougher bank of the flotation circuit in the plant.

Figure 5.5; shows MgO-nickel recovery curves after adding one Betamin depressant dose and that of adding two depressant doses. As a result of adding two depressant doses, there was a decrease in the MgO content in the successive concentrates that were collected during the flotation test as shown in Figure 5.5. The overall reduction in MgO content after adding the second depressant dose was 2.59%, while the nickel recovery was increased by 2.7%.

Thus, the results gave an indication that fast floating MgO silicates can depress the valuable mineral if the depressing effect of the depressant is short-lived.

This allowed fast floating MgO to float instead of the nickel sulphides and it explains why high MgO levels were obtained within the floatable particle size ranges in the rougher bank and final concentrate samples during the plant survey.

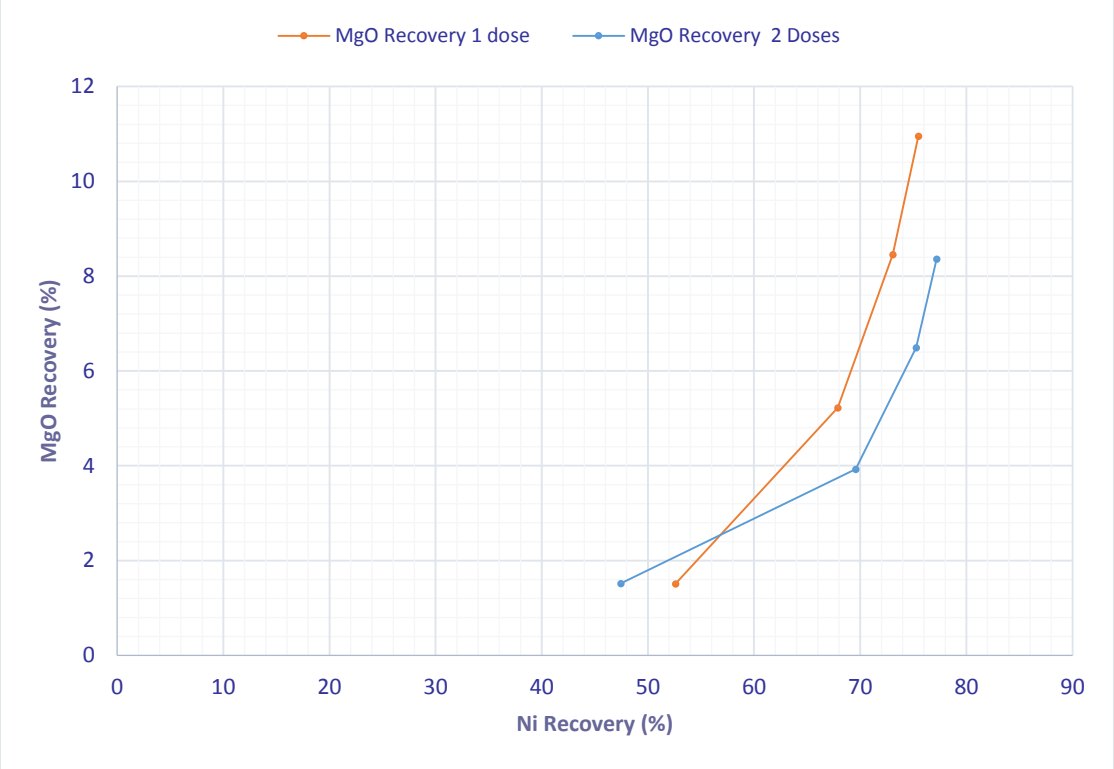

Figure 5.5 MgO-Nickel recovery curves for different depressant dosages

Figure 5.6; shows the MgO recovery-nickel grade curves for a single depressant dose and two depressant doses. The overall nickel grade of the concentrates were increased by 2.1%, after adding the second dose of the depressant.

Further supressing the fast floating MgO during the early stages of the flotation process in the rougher banks, by reducing the depressant dosage times from 10.4 to 5.2 minutes will improve nickel recoveries and grades. Hence, there is a need to change depressants addition points in the plant from the third tank to the second tank. Depressants are currently added in tanks 1 and 3 in the rougher banks. Depressant and other reagents point additions have been found to be also key in obtaining optimum results in the flotation circuits. This can be evaluated from the floatation kinetics results as well as the selectivity indices obtained from the batch floatation tests.

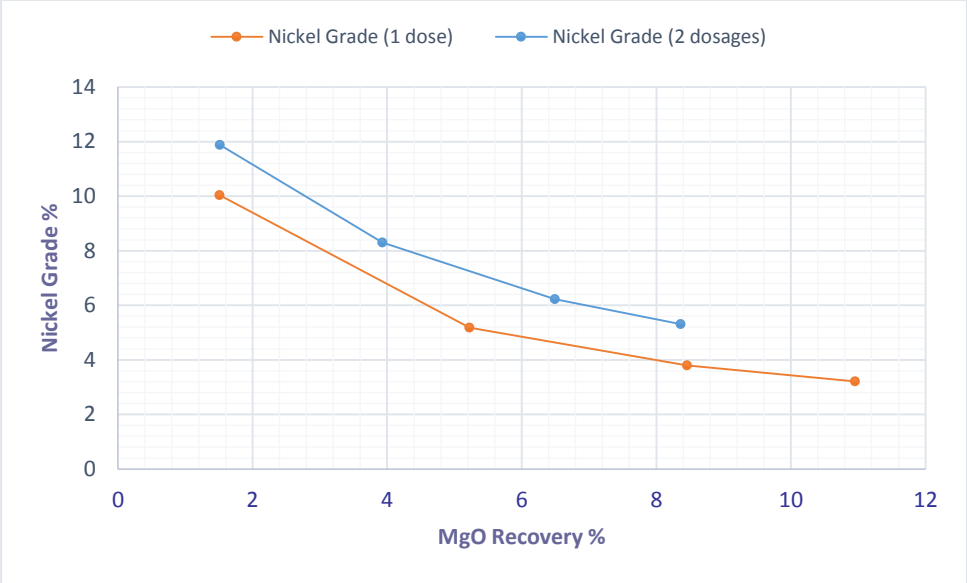

Figure 5.6 Nickel grade-MgO recovery curves for different depressant dosages

Reducing the residence time of the rougher banks could also result in the reduction of MgO content in the rougher bank concentrate. Most of the nickel was also recovered during the first five minutes of the test. The target recovery for the rougher bank is 60% and the test results show that at least 60% of the nickel was recovered during the first five minutes of the test. Thus, the residence time could be reduced without adversely affecting the nickel recoveries in the rougher bank.

5.2 Collector Combination Test Results

The collector combination results are shown in Table 5.4. PNBX, NC228 and NC236 collectors were blended with SIPX at a ratio of 1:1. SIPX collector had the highest selectivity index of 5.73 as compared to all the collector combinations that were tested. It had the highest nickel recovery that was 1.34% below the set target nickel recovery and its mass pull was 12.08%.

SIPX: NC236 and SIPX: NC228 combinations had fairly good selectivity indices; however, their nickel recoveries were 8.34% and 5.49% below the set target recoveries respectively. SIPX: PBNX had the lowest selectivity index of 5.31 and the lowest MgO recovery into the concentrate of 7.12%. Its nickel recovery was 3.95% below the set target recovery. Collector combinations are used in some plants in order to increase the selectivity of the reagent suite employed.

Table 5.4 Collector combination test results

Collector	SIPX only	SIPX and NC 228	SIPX and NC 236	SIPX and PNBX
Ni grade in feed %	0.87	0.86	0.93	0.76
Ni grade in concentrate %	5.3	5.3	5.21	5.26
Ni Recovery %	77.21	69.97	74.33	71.57
Ni grade in Tails %	0.26	0.27	0.27	0.29
MgO grade in Feed %	41.13	41.93	40.65	40.32
MgO Recovery in concentrate %	8.36	7.38	8.78	7.12
MgO grade in concentrate %	27.18	27.23	26.92	27.89
MgO grade in tails %	43.74	42.5	43.12	43.12
Mass of concentrate (g)	143.13	129.16	149.94	116.19
Selectivity Index	5.73	5.53	5.56	5.31
Ni Target recovery %	78.55	78.31	79.82	75.52
Variance	-1.34	-8.34	-5.49	-3.95
Mass Pull %	12.08	11.74	13.35	9.41
Enrichment Ratio	6.1	6.16	5.61	6.96

Different collectors have different properties and combining them has led to improved selectivity and flotation rates of different mineral processing plants.

Figure 5.7; shows the MgO recovery into the concentrates against the nickel recovery.

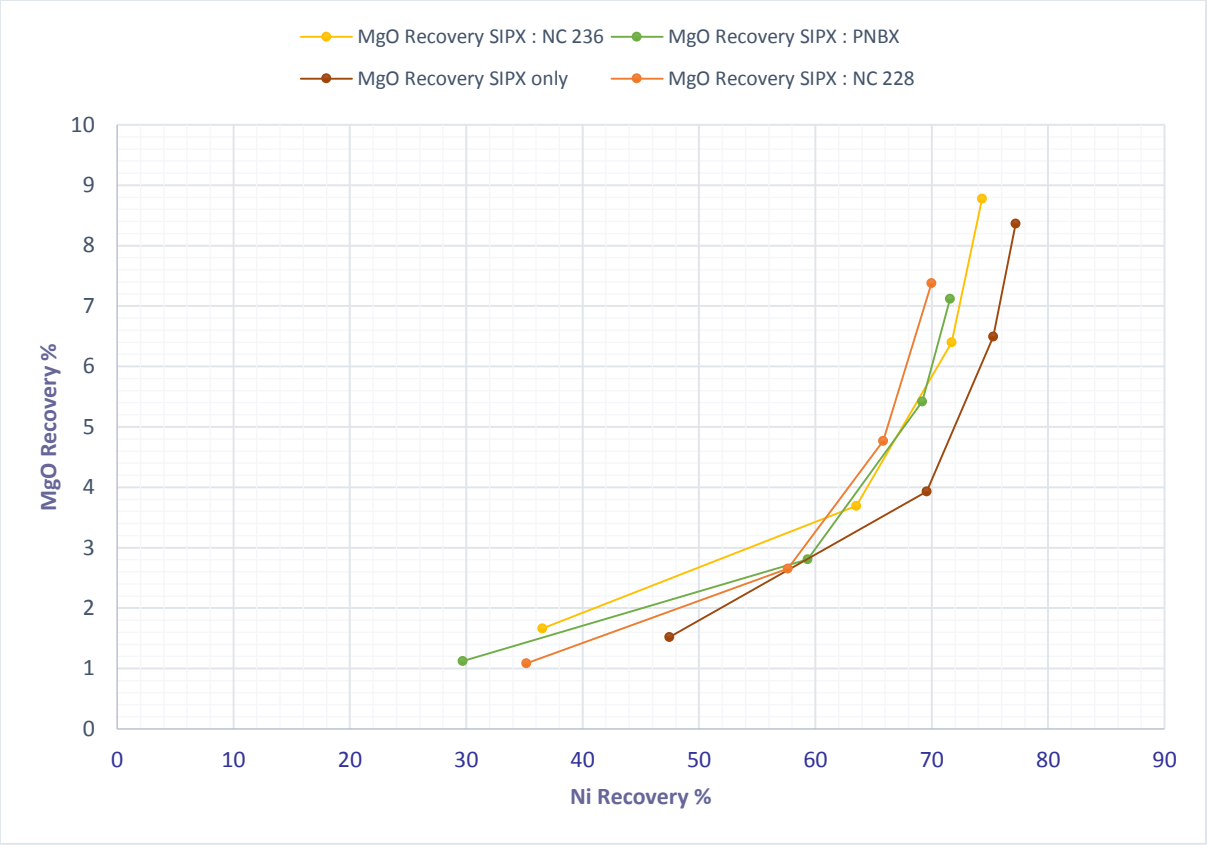

Figure 5.7 MgO-Nickel recovery curves for different collector combinations

Lower MgO recoveries also resulted in lower nickel recoveries in the SIPX: NC228 and SIPX: PNBX collector combinations. They had the MgO recoveries of 7.38% and 7.12% respectively. The collector combinations gave nickel recoveries that were below the set target and they had lower selectivity indices as compared to SIPX. They were more selective towards MgO as indicated by lower MgO recoveries into the concentrate, but there was no significant differences in the nickel grades obtained. These collector combinations were disqualified on the basis of their low nickel recoveries. The aim is to lower the MgO recovery into the concentrates without adversely reducing the nickel recoveries.

The SIPX collector that is currently used in the plant gave the best results, hence, there is no need to change the current collector. This collector will give better results when combined with a good depressants like the DLM RS and DLM PDE, which had high selectivity indices, lower mass pulls and long-lived depressing effect. These will make a better reagent suite as compared to the SIPX collector and the Betamin depressant that are currently used in the plant.

5.3 Effect of Rejecting Iron Sulphide from the Concentrates

The massive and disseminated ore was blended at a 1:1 ratio to give an average grade of 6.5% Ni, 26.5% Fe and 21.49% MgO. Figure 5.8; shows the nickel grade-recovery curve of the results obtained after carrying out the flotation tests with and without copper sulphate. A nickel maximum recovery of 95.81% and iron recovery of 82.67% were obtained after adding the copper sulphate activator. The overall concentrate grade obtained was 12.28% Ni and 43.0% Fe.

A nickel recovery of 90.07% and iron recovery of 27.56% were obtained without using a copper sulphate activator. The overall concentrate grade obtained was 20.54% Ni and 27.23% Fe. The mass of the concentrate obtained without using an activator was 315.84g while that obtained after adding an activator was 567.19g. This gave a reduction in the mass of the concentrates of 45.2% during the roughing stage and this will reduce the tonnage of concentrates that are shipped significantly.

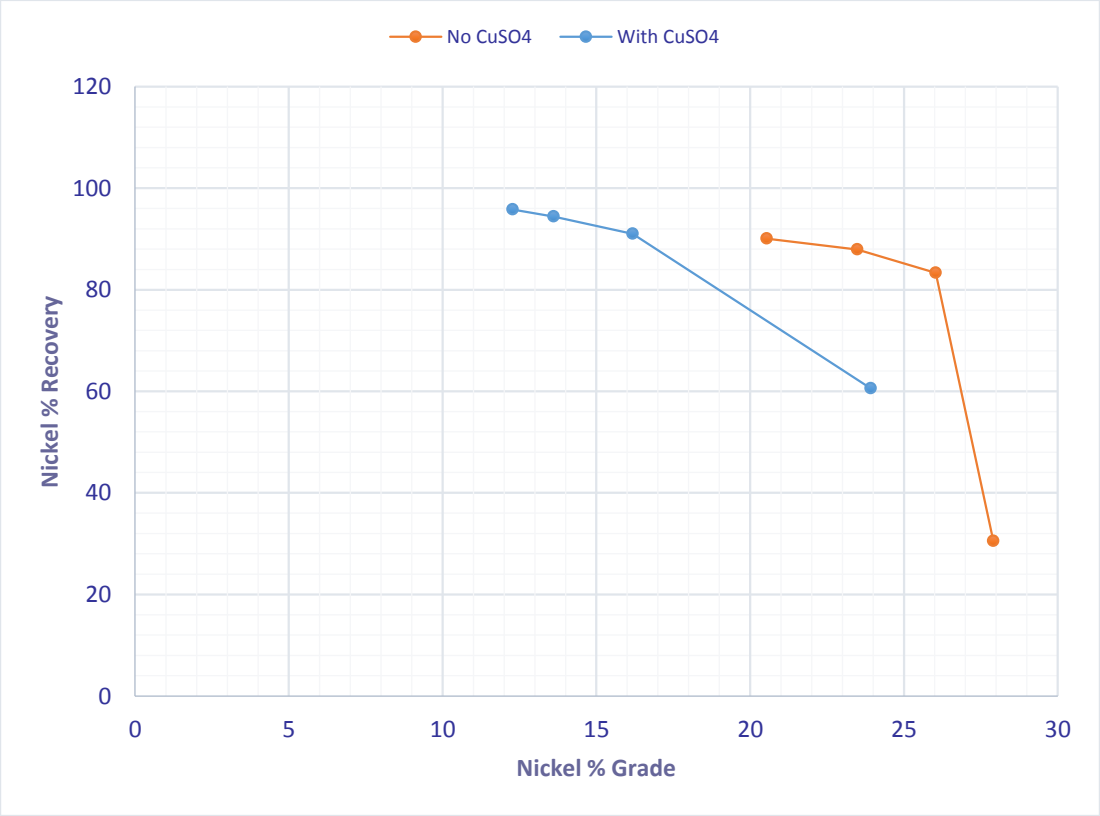

Figure 5.8 Nickel-grade recovery curves after adding an activator and without adding an activator

The iron-nickel recovery curve in Figure 5.9; showed that 56% of Fe was recovered into the concentrate during the first five minutes of the test after adding the activator as opposed to 19% recovered without adding the activator. This shows that most of the iron in the feed was fast floating. The last 26.67% recovered during the last 20 minutes of the test was slow floating as compared to the 8.56% Fe that was recovered without adding the copper sulphate. The recovery of iron into the concentrate was reduced by 55.1% by carrying out the test without an activator and the nickel recovery was reduced by 5%.

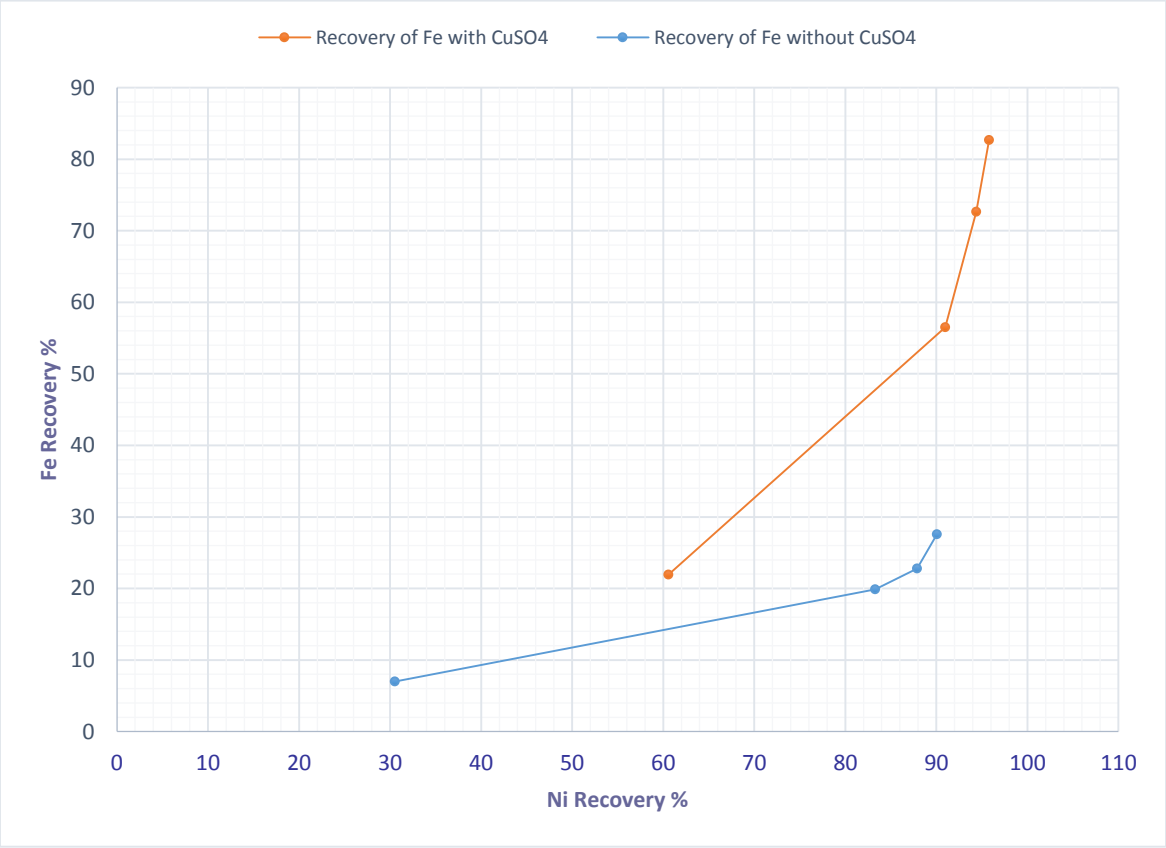

Figure 5.9 Iron-nickel recovery curves after adding an activator and without adding an activator

The copper sulphate activator is used in the bulk flotation of sulphides, which can also reduce the grade of the concentrates greatly if most of the nickel sulphides are barren. Some nickel processing plants rejects the nickel sulphides, especially when their nickel content is low and when further processing of the concentrates is carried out at a distant plant. Rejecting iron sulphides from the concentrate increased the concentrate grade and this can reduce shipping costs.

The recovery of iron into the concentrate was reduced by 55.1% by carrying out the test without an activator, while the nickel recovery was reduced by 5%. The nickel grade was increased by 8.26% after rejecting the iron sulphides. However, the reduction in nickel recovery into the concentrate needs to be weighed against the transportation costs. High iron recoveries into the concentrate when treating high grade ores ranges from 33%-40%. This results in reduced nickel grades in the concentrate, higher shipping costs and oxidation of the concentrates.

A plant trial for just a few hours should be carried out in order to accurately quantify the benefits of rejecting iron sulphide from the concentrates. An economic assessment should also be carried out in order to weigh the benefit of reduced shipping costs against the reduced nickel recovery. The total transport cost is currently at $102/tonne of concentrate as of 2016. Insurance and shipping costs amounts to 15% of the gross metal value. Rejecting iron sulphides will also enable the plant to operate at higher feed rates when milling high grade ores. This will reduce the mill residence times, power utilised, the production of excessive fines and nickel loses into the tailings. Chocking of the thickener pumps, caused by much higher mass pulls will also be eliminated.

CHAPTER 6

CONCLUSIONS AND RECOMMENDATIONS

6.1 Conclusion

Plant survey results showed that -32 µm particles had the highest recovery into the concentrates. High MgO contents were found in particle classes of 12.5-9.2µm in the final concentrate. The particle size range of 75µm to 32µm gave the good nickel grades and lower MgO contents in the final concentrate. Desliming the regrind mill feed led to reduced percentage of fines and MgO recovered into the scavenger final cleaner concentrate as compared to those recovered into the rougher final cleaner concentrate. The scavenger rougher bank feed, that was deslimed, had less finer MgO particles and MgO content as compared to the rougher bank feed. This shows that desliming before the coarse floatation process could lead to reduced MgO slimes in the feed and in turn reduces the reagent consumption in the bank. Introducing the desliming unit could be beneficial, since the desliming cyclones have low installation and operational costs. It will also reduce the recovery of MgO due to slime coatings in the final concentrates.

MODSIM simulation results showed that overgrinding in the primary milling circuit is most likely to take place when the feed is blended with the higher percentage of massive ores, whose milling rate was found to be three times higher than that of disseminated ores or due to the high percentage of fines in the circulating load due to excessive spigot wear when milling disseminated ores. Hence, there is a need to reduce the mill residence times when milling the high grade ores. Optimum milling conditions still need to be determined using a simulator for a range of high grade feed assays that range from 1.7% -4% Ni. Using a professional simulator that allows the variation of more parameters will enable the attainment of more accurate results as compared to the MODSIM academic version.

The DLM RS depressant gave the best results, hence, a plant trial should be carried out with the depressant. It gave an overall reduction in the MgO content in the concentrates of 3.79% and it reduced the mass of concentrates collected by 32% as compared to the Betamin depressant.

It had higher selectivity towards MgO silicates and also had higher recoveries as compared to the Betamin standard depressant that is currently used in the plant.

Carrying out a test without a depressant showed that 57.7% of the MgO was recovered during the first five minutes of the test. The highest MgO content was recovered during the first minute, indicating that most of the MgO was fast floating. Hence, there is a great need to effectively supress the fast floating MgO during the early stages of the flotation process.

Further supressing the fast floating magnesium silicates by adding a second depressant dose after one minute into the flotation test increased the nickel grade by 2.1% and the nickel recovery by 2.7%. This shows that the MgO silicates could depress nickel when the depressing effect of the depressant used is short-lived. Reducing the depressant dosage times in the plant from 10.4 to 5.2 minutes will improve nickel recoveries and grades. It was also noted that 60% Ni was recovered during the first five minutes of the test. Therefore, reducing the residence time of the rougher banks could also help reduce the amount of MgO recovered into the rougher concentrates without any adverse effects on nickel recovery.

The SIPX collector, which is currently used in the plant showed better selectivity towards MgO as compared to the other collector combinations that were tested. Its combination with depressants that have greater selectivity towards MgO and long-lived depressing effect such as DLM RS and DLM PDE gave better results. This is a better reagent suite as compared to the SIPX collector and Betamin depressant that is currently used in the plant.

Recovery of iron and nickel into the concentrate were reduced by 55.1% and 5% respectively by carrying out the flotation test without an activator. The concentrate grade was increased by 8.26% and the mass of the concentrate recovered was reduced by 45%. A plant trial, however, should be carried out for a few hours to accurately quantify the benefits of rejecting iron sulphide from the concentrates. The test was carried out using a sample that had higher nickel and iron contents as compared to the usual plant grades.

6.2 Recommendations

- A plant trial using the DLM RS depressant, which had better selectivity towards MgO and higher nickel recoveries as compared to the Betamin standard depressant that is currently used in the plant should be carried out.
- It is recommended that the rougher bank residence time be reduced and the second depressant dose be added after 5.2 minutes instead of 10.4 minutes to effectively supress the fast floating MgO silicates.
- A plant trial should also be carried out for a few hours, floating without copper sulphate, to accurately quantify the benefits of rejecting iron from the concentrates.
- To reduce overgrinding, it is recommended to increase the feed rate of the high grade ores and also reject iron from the concentrates to avoid chocking of the thickener pumps.
- Simulation tests also need to be carried out on different blending ratios in the plant. Blending ratios should be pre-determined in order to avoid the wide variations in the feed grade.
- Worn out cyclone spigots must be replaced promptly to reduce the percentage of fines in the circulating load.

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8 APPENDIX

Appendix A: Simulation data and Results

Table A1: Mill 1 and 2 feed PSD

Sieve size in microns	Mass retained in grams	Cumulative Passing mass	% Passing
19 050	41.19	41.19	98.91
16 000	113.85	155.04	95.91
9 050	1578.1	1733.14	54.29
4 000	1058.74	2791.88	26.37
3 350	93.60	2885.48	23.90
1 700	274.7	3160.18	16.66
1 180	85.43	3245.61	14.40
850	64.08	3309.69	12.71
600	60.64	3370.33	11.11
425	51.04	3421.37	9.77
300	50.36	3471.73	8.44
212	50.54	3522.27	7.11
150	47.2	3569.47	5.86
106	47.44	3616.91	4.61
75	43.77	3660.68	3.46
45	32.88	3693.56	2.59
-45	98.18	3791.74	0.00
	3791.74		

Table A2: PSD after milling low grade ore for 0.5, 1, 2 and 4 minutes

Screen size mm	Size class	0 minutes	% mass retained	0.5 minutes	% mass retained	1 minute	% mass retained	2 minutes	% mass retained	4 minutes	% mass retained
1.7	1	34.18	11.393	26.4	8.7397	24.86	8.24025	17.65	5.864567	12.44	4.153728
1.18	2	63.37	21.1226	59.2	19.5981	53.52	17.7401	56.69	18.83639	42.76	14.27761
0.85	3	43.37	14.4562	43.53	14.4106	39.69	13.1559	43.03	14.29758	37.91	12.65819
0.6	4	38.92	12.9729	40.08	13.2684	41.62	13.7956	41.2	13.68953	38.49	12.85185
0.425	5	32.08	10.693	33.94	11.2358	33.84	11.2168	34.4	11.43009	35.45	11.83679
0.3	6	26.61	8.8697	27.99	9.26606	29.66	9.83128	29.74	9.881712	31.6	10.55127
0.212	7	22.24	7.41309	23.49	7.77634	25.87	8.57503	24.66	8.19378	27.83	9.292464
0.15	8	18.88	6.29312	19.75	6.53822	21.84	7.23922	20.88	6.937799	23.96	8.000267
0.106	9	16.92	5.63981	19.91	6.59119	19.13	6.34095	17.08	5.675173	21.72	7.252329
-0.106		3.44	1.14663	7.78	2.57556	11.66	3.86489	15.63	5.193381	27.33	9.125513
		300.01		302.07		301.69		300.96		299.49	

Table A3: PSD after milling the massive ore for 0.5, 1 and 2 minutes

Screen size mm	Size class	0 minutes	% mass retained	0.5 minutes	% mass retained	1 minute	% mass retained	2 minutes	% mass retained
1.7	1	38.84	12.8981	15.76	5.23762	10.54	3.50108	4.77	1.580988
1.18	2	71.63	23.7871	45.22	15.0282	32.21	10.6992	16.1	5.33625
0.85	3	54.31	18.0354	46.19	15.3506	37.35	12.4066	20.78	6.887408
0.6	4	59.36	19.7124	63.64	21.1499	60.11	19.9668	41.16	13.64224
0.425	5	55.35	18.3808	66.02	21.9408	67.36	22.375	64.25	21.29528
0.3	6	19.72	6.54867	33.96	11.2861	41.11	13.6555	55.68	18.45481
-0.3	7	1.92	0.6376	30.11	10.0066	52.37	17.3958	98.97	32.80302
		301.13		300.9		301.05		301.71	

Scale-up of lab batch data for full-scale continuous mill simulation

$$S_i(d) = a_T \left(\frac{X_i}{X_0}\right)^{\alpha} \left(\frac{1}{1+\left(\frac{X_i}{C_1 \mu_T}\right)^{\wedge}}\right) C_2 C_3 C_4 C_5 \ldots\ldots (11)$$

Where

$$C_1 = \left(\frac{D}{D_T}\right)^{N_2} \left(\frac{d}{d_T}\right)^2$$

$$C_2 = \left(\frac{d_T}{d}\right)^{N_0}$$

$$C_3 = \left(\frac{D}{D_T}\right)^{N_1}, \text{ D} \leq 3.81\text{m}$$

$$= \left(\frac{3.81}{D_T}\right)^{N_1} \left(\frac{D}{3.81}\right)^{N_1 - \Delta}, \text{ D} \geq 3.81\text{m}$$

$$C_4 = \left(\frac{1 + 6.6{J_T}^{2.3}}{1 + 6.6{J_T}^{2.3}}\right) exp[-c(U - U_T)]$$

$$C_5 = \left(\frac{\phi_c - 0.1}{\phi_{c_T} - 0.1}\right)\left(\frac{1 + \exp[15.7(\phi_{c_T} - 0.94)]}{1 + \exp[15.7(\phi_c - 0.94)}\right)$$

D = mill diameter, m

d = ball diameter, m

X = particle size, mm

X_0 = 1 mm;

Subscript T refers to test mill

J = fractional load volume

U = fractional void filling

ϕ_c = mill speed fraction of critical speed

C_1 = adjust peak position for S_i

Other C's adjust Si values

N_0 = 1; N_1= 0.5; N_2= 0.2; Δ= 0.2; c= 1.32

A and μ do not change on scale-up

Table A4: Scale up for ball mill 1

SCALE UP FOR MILL 1	
ϕ mill	0.7149
ϕ test mill	0.75
D test mill	0.203
D mill	3.66
d mill	0.0785
d test mill	0.03
J test	0.27
J	0.4
U	1
C1	12.2093857
C2	0.382165605
C3	4.246122543
C4	0.882292199
C5	0.965720672
a test	0.348214659
μ test	2.36
S (1 mm)	0.240681085

Table A5: Scale up for ball mill 2

SCALE UP FOR MILL 2	
ϕ mill	0.7376
ϕ test mill	0.75
D test mill	0.203
D mill	3.66
d mill	0.0785
d test mill	0.03
J test	0.27
J	0.4
U	1
C1	12.2093857
C2	0.382165605
C3	4.246122543
C4	0.882292199
C5	1.001371769
a test	0.348214659
μ test	2.36
S (1 mm)	0.249566206

Table A6: Selection function for each screen size

Microns	Selection function
2360	0.412689339
1700	0.293459157
1180	0.208675797
850	0.149619075
600	0.10639259
425	0.07565468
300	0.053734964
212	0.038210382
150	0.027171011
106	0

Table A7: Breakage function results

Microns	Breakage Function									
2360	0	0	0	0	0	0	0	0	0	0
1700	0.379056613	0	0	0	0	0	0	0	0	0
1180	0.114007628	0.364646	0	0	0	0	0	0	0	0
850	0.098228814	0.123122	0.372567	0	0	0	0	0	0	0
600	0.078487997	0.098369	0.120497	0.37122	0	0	0	0	0	0
425	0.063986442	0.080194	0.098229	0.121844	0.372567	0	0	0	0	0
300	0.051440357	0.06447	0.078969	0.097948	0.121235	0.372136	0	0	0	0
212	0.041367773	0.051846	0.063506	0.078769	0.097491	0.120929	0.371652	0	0	0
150	0.033508349	0.041996	0.05144	0.063804	0.078969	0.097948	0.121413	0.372136	0	0
106	0.139916028	0.175357	0.214792	0.266415	0.329738	0.408988	0.506936	0.627864	1	0

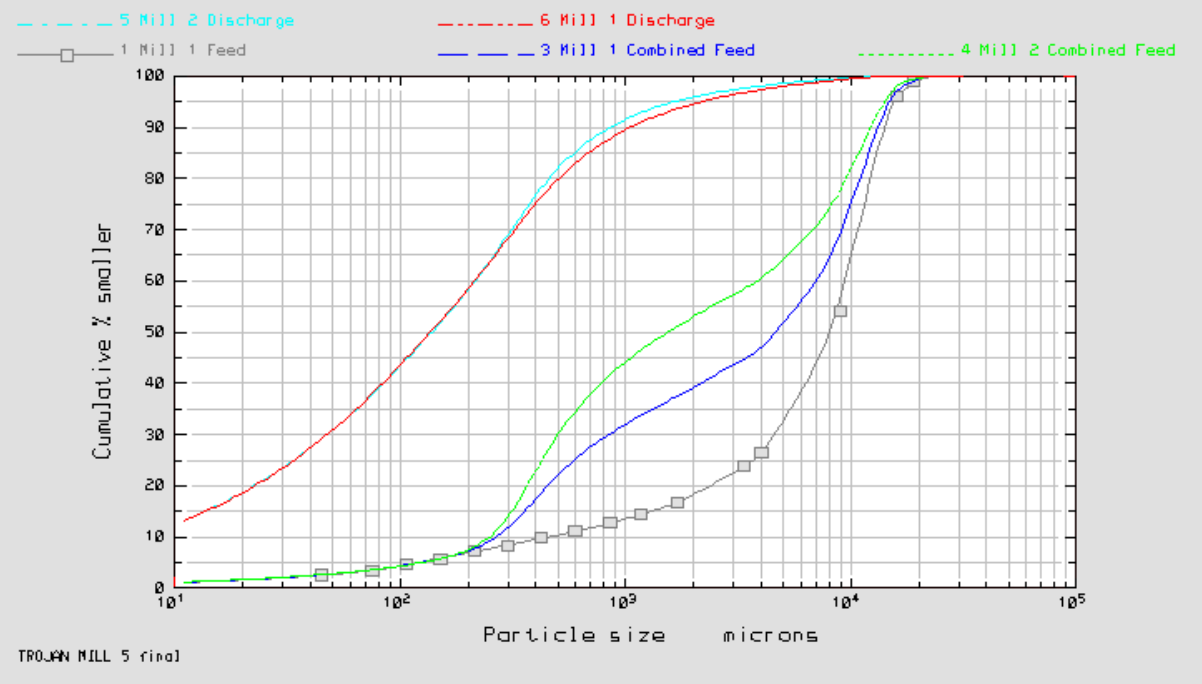

Figure A1: Mills PSD graphs from the MODSIM simulation results

Appendix B: Flotation Test Results

Table B1: Cumulative recovery of Guargum depressants

Ni % Recovery at t minutes	1	5	15	25
Betamin (std)	46.8	66.2	71.6	74.5
Betamin with NaCO3	52.6	67.9	73.1	75.5
No Depressant	50.7	75.3	83.6	86.9
Cytec S9349	33.3	55.2	63.7	67.0
Cytec S9349 With NaCO3	41.9	60.2	66.7	69.9
DLM PDE	33.7	61.7	71.8	74.9
DLM PDE With NaCO3	39.3	58.1	64.9	68.4
DLM RS	58.4	71.6	76.7	78.4
DLM RS With NaCO3	44.7	62.1	67.3	69.2

Table B2: Cumulative recovery of CMC depressants

Ni Recovery at t minutes	1	5	15	25
Depramin 267	51.7	63.5	68.1	70.7
Depramin 347	45.3	60.7	63.9	66.2
Depramin 177	47.2	61.3	66.7	69.7
ND 523	51.1	63.6	68.5	71.1
ND 522	54.7	67.4	71.9	74.5
ND 521	43.9	54.5	59.1	61.6
ND 522 (normal pH)	41.6	57.0	60.6	62.2
Depramin 267 (normal pH)	57.4	71.3	76.0	77.5